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DISPOSAL OF BRINES PRODUCED IN RENOVATION OF MUNICIPAL WASTEWATER

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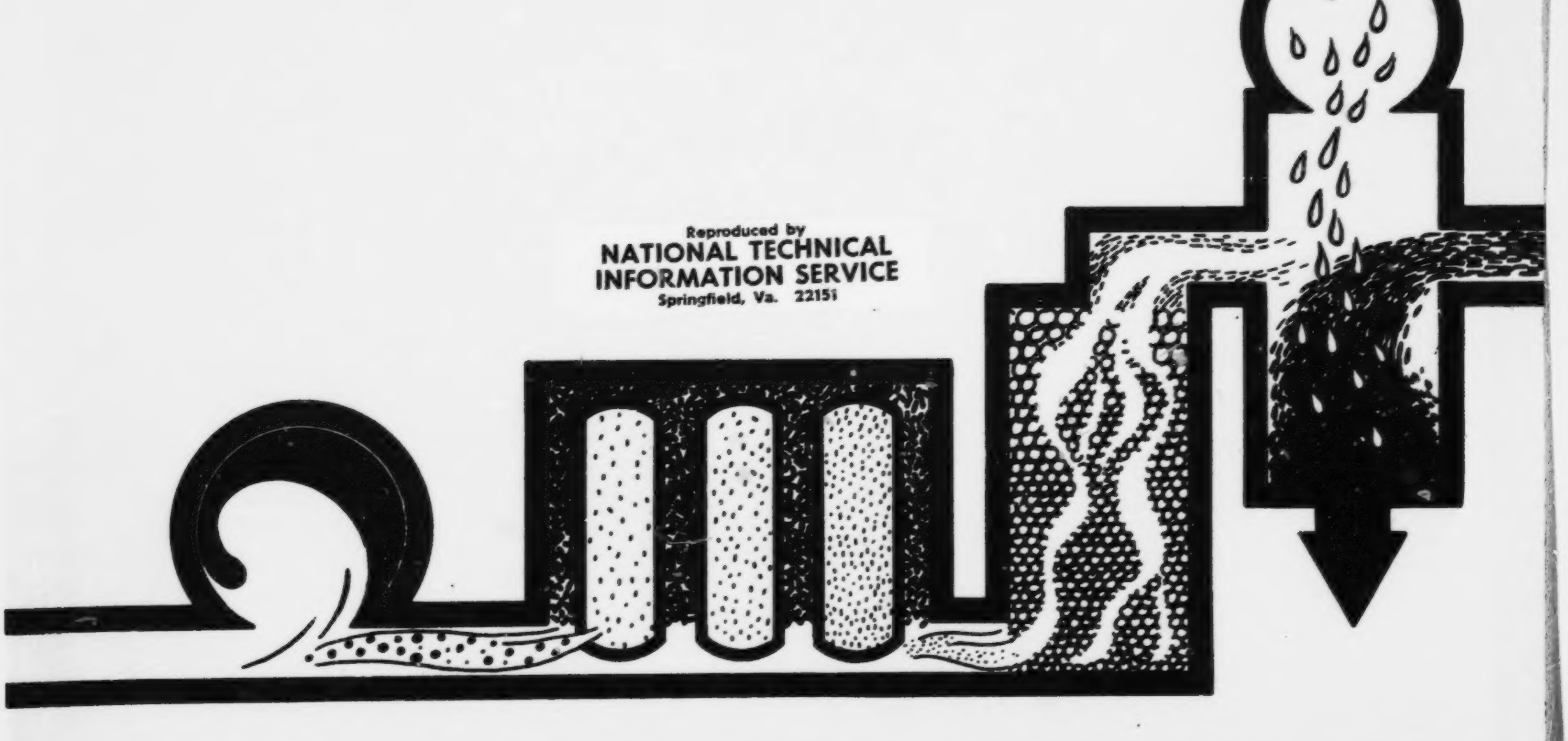

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27 Abstract

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Near El Paso, Texas, brine can be dumped on worthless arid land at a cost of $.052/Kgal. It can be injected into the saline Hueco-Bolson Basin at $0.13/Kgal. Solar evaporation in local ponds, using 30 mil liners and a pipeline to convey residual brine 50 miles for ultimate disposal, costs $0.18 Kgal.

Solar evaporation east of Denver, using ponds with a 30 mil liner, would cost $0.76/Kgal. Alternately, solar evaporation east of Pueblo, Colorado in lined ponds would cost $0.96/Kgal., including the pipeline from Denver. Multistage flash evaporation to 10% solids would reduce the amount of brine and the size of the solar ponds to a point where they might be acceptable. Their combined cost, based on $0.46/mbtu steam and steam-driven pumps is $0.54/Kgal. of brine effluent. Well injection is unfeasible here, due to earthquakes.

At Tucson, the temporary measure of using injection wells to 3500 feet while awaiting the Southwest Water Plan would cost $0.13/Kgal. A permanent scheme, using local solar ponds with 30 mil liners would cost $0.18/Kgal., including costs for a residual brine pipeline to the Wilcox Plaza 50 miles eastward.

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DISPOSAL OF BRINES PRODUCED IN RENOVATION OF MUNICIPAL WASTEWATER

by

Burns and Roe, Inc.
Oradell, New Jersey 07649

for the

FEDERAL WATER QUALITY ADMINISTRATION

DEPARTMENT OF THE INTERIOR

Program #17070 DLY
Contract #14-12-492
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Cincinnati, Ohio
May, 1970

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ABSTRACT

Costs of ultimate disposal of brine wastes from municipal water renovation schemes have been investigated for the sites of El Paso, Texas Tucson, Arizona and Denver, Colorado. Based on 10 million gallons per day, 7% fixed charge rate, and 12 mills/Kwhr power cost, estimated costs are as follows:

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This report was submitted in fulfillment of Contract 14-12-492 between the Federal Water Pollution Control Administration and Burns and Roe, Inc., (Program #17070 DLY).

Key Words: ultimate disposal, brine wastes, municipal renovation schemes, deep well injection, solar evaporation, brine reduction, flash evaporation, disposal costs.

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INTRODUCTION

AUTHORIZATION

On February 25, 1969 Burns and Roe, Inc. was authorized by the United States Department of the Interior, Federal Water Pollution Control Administration - under Contract No. 14-12-492 - to study economical solutions to the problem of ultimate disposal of brine wastes from advanced waste treatment processes.

Advanced waste treatment is the processing of waste waters beyond that required for pollution control to permit beneficial reuse. A major cost factor associated with advanced waste treatment is the disposal of brines resulting from these treatment processes.

Without ultimate disposal, the pollution problems are simply transferred from one locality to another. With it, the brines are ultimately disposed of in as nearly a non-polluting fashion as possible; for example, to the oceans, to existing salt beds, or to underground saline waters.

OBJECTIVE

This investigation has been undertaken for the FWPCA to determine the cost of brine (liquid-waste) disposal facilities for several sites in the United States.

The methods of disposal are themselves combinations of one or more operations. The costs of these separate operations will be determined at the several sites to be investigated. Total costs of the most economical combined disposal schemes will be determined as a function of brine concentration and quantity.

SCOPE OF WORK

Review the state-of-the-art for feasible methods of concentrating and disposing of brine wastes. Process design and cost figures will be obtained or developed for operations such as deep-well disposal, concentration to saturation or dryness by submerged combustion, multiple-effect evaporation and/or solar evaporation, hauling of dry salt, pipeline transport, and storage pond impoundment with periodic discharge.

Determine the costs of these separate operations at three separate locations in the United States to be established in later discussions with officials of FWPCA. Costs are to be determined as a function of initial brine concentration, final brine concentration and total brine throughput. Emphasis will be placed on obtaining accurate cost figures.

Determine the total cost of disposal of brine by several of the most promising combinations of the separate operations for which costs have been obtained. Typical schemes may be:

Concentration of brines to saturation followed by pipelining to the nearest ocean or salt lake

Concentration of brines to dryness and hauling of dried salt to nearest salt bed

Deep-well disposal with or without prior concentration

Impoundment of brine wastes with controlled disposition during spring floods

The cost of the most promising schemes will be determined as a function of the total amount of brine solids for which disposal is required and the concentration of the brine. The amount of brine solids handled at each site will be computed as that from municipal wastewater renovation plants of 1 MGD, 10 MGD and 100 MGD capacity. The initial brine concentration for each of the sites studied will be selected as the most promising value determined by reviewing the state-of-the-art and by consulting with FWPCA and with vendors for water and wastewater renovation processes currently proposed (such as electrodialysis and reverse osmosis).

The influence of variations in brine composition on the cost of ultimate disposal will be predicted on the basis of current knowledge of these variations. Equipment, materials, processing and disposal problems uncovered during this study will be reviewed. Recommendations will be made as needed for research and development work in these areas.

TECHNICAL APPROACH

Item 1 of the Scope, in which the word "operations" is followed by a list of possible processes to be studied, allows certain of the processes which appear to be least feasible to be given merely cursory treatment, while others that were not included in the original list can be covered in some detail. Thus, storage pond impoundment with periodic discharge, barium-zeolite desulfation, and multistage flash evaporation using an immiscible heat exchange liquid, have not been covered in great detail, since they do not appear to be among the "feasible methods" applicable to concentrating and disposing of brine wastes under the ground rules of this study.

On the other hand, additional concentration methods have been added so as to expand the list of item 1. Electrodialysis with ion-specific membranes to allow concentration without scaling, and submerged combustion were included among the alternative methods studied. In addition, the use of vapor-compression evaporators to concentrate all the way to saturation has been analyzed, and parametric unit cost and operating cost equations have been developed.

Item 3 of the Scope lists typical disposal schemes that "may be" considered. Because "disposal during spring floods" does not appear applicable to the selected sites, it was not considered under Item 3. Additional work under this item includes examination of pretreatment and post-treatment of AWT effluents, to render the effluent suitable for standard quaternary treatment. Electrodialysis, Reverse-Osmosis, and Ion Exchange were the wastewater renovation methods chosen. Particular attention under Item 3 has been devoted to concentration by conventional multistage flash evaporation as the best alternative to solar evaporation up to 10% total dissolved solids (tds).

A final selection of sites for the brine disposal study has been made on the following basis:

1. El Paso, Texas; population 300,000. Needs water. Location, hot, dry, on Rio Grande River with highly variable flow, used for irrigation downstream.

2. Tucson, Arizona; population 250,000. Needs water. Location, hot dry, inland. Probably ideal site for solar evaporation ponds. Salt pollution of Gila River and tributaries is serious.

3. Denver, Colorado; population 520,000. Seeks maximum water reuse to meet anticipated growth. Location, cool, dry, inland on South Platte River. Deep well disposal of waste brines not feasible due to earthquakes. Suggests creation of a salt water recreational lake as a possible disposal method.

SUMMARY AND RECOMMENDATIONS

SUMMARY

The following Summary Sheet (Table 1) lists disposal costs in \$/Kgal. of brine for 3 sizes of disposal plants based on 5% and 10% annual fixed charge rates at the 3 selected sites. For additional information on these methods, see the individual sections of the report.

RECOMMENDATIONS

El Paso, Texas

A general solution for the water problems of the region must await implementation of the Texas Water Plan, whereby storage basins now being exhausted can be replenished and stabilized with Mississippi River water. As the best temporary expedient until then for the City of El Paso, injection of brine into the East end of the Hueco Bolson Basin at 3500 ft. is recommended.

Denver, Colorado

Water here is of pristine quality, and a multiple reuse scheme of 197 mgd is considered only from a hypothetical standpoint. This reuse scheme, which would satisfy city needs beyond 1985 with no additional water withdrawal from the Colorado River, would employ multistage flash preconcentration to 10% solids, with 0.7 mgd of evaporator blowdown going to solar ponds East of the city for ultimate disposal.

Tucson, Arizona

The need for a reuse scheme here will require the exportation by 1985 of all salt entering the Tucson Basin in amounts equal to the salt inflow due to importing water from the Parker Dam on the Colorado River. The most economical scheme will be to provide local solar evaporation ponds, with the blowdown from these ponds being pipelined 50 miles East to the closed and already contaminated Wilcox Playa Basin.

FUTURE STUDY

Means whereby the tds removal step of an advanced waste treatment plant can be made to concentrate the brine as well as purify the secondary sewage effluent should be investigated. These are outside the scope of the present study. Such means include:

Selective electrodialysis whereby hardness ions, necessary for non-aggressive product water, are left in the dialyzate, while monovalent cations, such as $NH_4{}^+$, Na^+ and K^+ are concentrated to saturation for ultimate disposal. For typical municipal effluents, combinations of pretreatments and post-treatments and ion specific membranes would have to be investigated.

Multistage flash evaporators optimized for concentrating secondary sewage effluent to 10% solids, with lime softening and ion exchange as a pretreatment, and carbon adsorption as a post-treatment. Brine carryover, ammonia carryover, condenser leakage, biological contamination and odor would be problems that would have to be solved by suitable post-treatments to render the product water fit for reuse.

Combinations of the above processes with reverse osmosis, using in situ casting and also the new high-efficiency cellulose-acetate-butyrate membranes. It would seem possible to tailor the tds removal and concentrating step to the requirements of any given municipality.

Special handling of cooling tower blowdowns and other unusual industrial wastes that presently wind up in municipal effluent streams.

TABLE 1

SUMMARY SHEET

COMPARATIVE COSTS OF BRINE DISPOSAL AT THREE SITES

Brine Quantity (mgd)	0.1	1.0	10.0	0.1	1.0	10.0
Fixed Charge Rate (%)		10			5	
	Costs ($/Kgal.)			Costs ($/Kgal.)		
EL PASO						
Dumping at White Sands	0.47	0.19	0.07	0.27	0.10	0.04
Deep Well in Hueco Basin	0.28	0.16	0.16	0.21	0.12	0.11
DENVER						
Lined ponds (30 mils) at Denver	1.26	1.125	1.08	0.63	0.56	0.54
Lined ponds (10 mils) at Denver	0.678	0.54	0.498	0.34	0.27	0.25
Pipelining to Pueblo plus (30 mils) ponding at Pueblo	4.18	2.16	1.35	2.33	1.15	0.69
Pipelining to Pueblo plus (10 mils) ponding at Pueblo	3.71	1.7	0.88	2.09	0.91	0.46
Forced evaporation plus (30 mils) ponds	1.07	0.84	0.69	0.75	0.61	0.51
Forced evaporation plus (10 mils) ponds	1.02	0.78	0.63	0.72	0.58	0.48
Deep Well	Not Possible			Not Possible		
TUCSON						
Lined ponds (30 mils) plus pipelining of brine concentrate	1.37	0.52	0.25	0.70	0.27	0.14
Lined ponds (10 mils) plus pipelining of brine concentrate	1.20	0.35	0.20	0.61	0.18	0.10
Deep Well	0.28	0.16	0.16	0.21	0.12	0.11

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GENERAL BACKGROUND

SALT BUILDUP PROBLEM

Many areas in the western portion of the United States are characterized by arid or semi-arid conditions, with resulting shortages of water supply. As a consequence, water is too valuable a commodity to be used once and then thrown away. It must be reused. The reuse may be applied to agricultural, municipal, or industrial requirements, depending on the ease of treatment. In that area of the country, reuse has primarily been applied to irrigation for agricultural purposes. Since with such reuse, there is insufficient water to provide suitable irrigation return flows, or net annual runoffs from the irrigated areas, soluble soil solids are continually leached to the surface and concentrated there along with the solids of the irrigation water. The net consequence of this procedure is the destruction of the irrigated land, usually by buildup of sodium salts at the surface.

In a typical United States community, reuse of the municipal effluent can be expected to add 350 mg/l total dissolved solids (tds) to the water on each cycle. If the average water supply contains over 150 mg/l tds, it follows that a single reuse will raise the tds to over 500 mg/l, the potability limit recommended by the U.S. Public Health Service, and hence render the water unfit for municipal reuse. The problem is made more acute by the nature of industrial pollutants, accounting for about 150 mg/l of the 350 mg/l added per use, and frequently containing toxic as well as refractory substances. Pesticide residues, photographic wastes, chlorinated phenols, hexavalent chromium, copper and boron compounds are among such objectionable substances and are not removable by conventional primary and secondary waste treatments. These and many other refractory substances occur in the municipal wastes from heavily industrialized areas and must be removed before the water can be considered fit for reuse.

WATER RENOVATION TO REMOVE SALTS

The above problems due to salt buildup can usually be solved by one of the modern desalination methods operating on the municipal waste water. For water sources below 2,000 mg/l, three processes show good potential for desalination: ion-exchange, reverse osmosis, and electrodialysis. Ion exchange in combination with chemical processes has the dual potential of producing potable water while at the same time concentrating the solids up to 8% by weight. It has the disadvantage of contributing solids to the salt disposal problem. Reverse osmosis has the advantages of being a nonselective, highly dependable, low-operating-cost process. It has the disadvantages of discharging large volumes of low concentration brine, from which the pump work must be extracted by turbines for good power economy. Electrodialysis has the advantages of being fully commercial, having good power economy, and also having a good concentrating factor for the waste brine. Its main disadvantages are membrane poisoning, polarization, and exfoliation of the membranes by scale when improperly operated. In common with distillation and freezing processes, the above three processes produce large volumes of waste brine, containing all of the contaminants removed in rendering the municipal effluent suitable for multiple reuse. Determination of the costs for ultimate disposal of these waste brines is the main objective of this study.

BRINE DISPOSAL MEANS

The amount of waste brine blowdown in a desalting operation is considerable. For example:

- A 100-mgd water renovation plant produces a blowdown of 10 mgd of brine of about 7000 mg/l concentration. Disposal of this quantity of brine is a problem of no small magnitude.
- If concentrated to 30 percent solids, the volume of brine waste is still considerable - 180,000 gallons per day - and it would still contain all of the harmful salts.
- Even as dry salt with a bulk density of about 125 lb/ft^3, the daily production would occupy almost 5000 cubic feet.

Koenig has considered the following brine disposal alternatives (Koenig, '58, DePuy, '69):

Disposal may be accomplished by underground injection, land dumping on unusable and worthless areas, sea discharge via pipeline, stream discharge, or abandonment at the operation site. An intermediate process of evaporation to saturation or dryness, and conveyance operations may also be used in combination with the final disposal operations.

Injection disposal involves a comprehensive geologic investigation and field testing of the disposal zone and reservoir areas to determine that safe, effective underground disposal is possible, that the injected brine is compatible with fluids in the reservoir, that it cannot encroach on or pollute underground fresh water, that future natural resources will not be contaminated, and that the risk of causing unforeseen phenomena, such as earthquakes and eruptions, is minimized.

Land dumping is feasible in certain western locations, such as remote deserts and playas having no net annual runoff and no useful aquifers beneath, or occupying closed basins where the land has already been rendered useless for agricultural or other human purposes.

Conveyance via pipeline to the sea or to a salt lake basin is feasible for those inland sites fortunate enough to be located within 150 miles of such an ultimate disposal sump. Outfalls would have to be located sufficiently offshore out into an ocean current, so that natural mixing and dispersion could be relied upon to prevent ecological damage.

Stream discharge is generally not feasible in the United States, and because of the local pollution problem that it introduces, it does not really qualify as an ultimate disposal method. By impoundment of brines with controlled release during floods, however, the pollution problem can be more or less equalized throughout the year and the periods during which legal limits on pollution are exceeded can be minimized.

Abandonment at the site of evaporation is feasible in the far west, where the annual gross evaporation rate greatly exceeds the annual rainfall. The salt residues in a lined evaporation pond remain in the pond and are left to build up continuously over the lifetime of the desalination plant. Alternately, lined solar evaporation ponds may be used to preconcentrate to saturation near the desalination facility, and the concentrated brine from this can be dumped on useless land in a remote area (DePuy, '69).

Preconcentration techniques, whereby the brine is concentrated to saturation or dryness before conveyance to the ultimate disposal site include both chemical and physical methods. The successful application of one or more of these methods in geographically feasible areas will depend entirely upon the economics of the local situation. Lime coagulation, precipitation, and ion exchange, are frequently inexpensive chemical methods. Solar evaporation, multistage flash evaporators, and vertical tube evaporators are effective preconcentrating devices for concentrating waste brines to the point where they can be most economically disposed of as a liquid, a thick slurry, or a solid. For disposal as a liquid, part of the wastes is evaporated, and the remaining part is pipelined away to the nearest non-leaching sump or ocean. For disposal as a slurry, the initial waste brine is concentrated to the point at which scaling or crystallization occurs. For disposal as a solid, the initial liquid is concentrated to saturation and then dried to solids in solar evaporation ponds.

While the costs of solar evaporation in arid areas are always favorable, the costs for forced evaporation by mechanical means can be expected to be high, particularly when the added costs for crystallizing and for drying to solids are considered. There are several broad categories of equipment that have been proved commercially:

> To concentrate up to but not exceeding 10% solids, use of the conventional single-effect multistage flash-evaporator that has already been developed for seawater conversion results in a low cost per thousand gallons of water evaporated. Use of these units on wastewater instead of seawater, however, requires a reoptimization to determine minimum costs since the previous results for seawater are not directly applicable.
>
> For concentrations from 10% to 40% total dissolved solids, evaporation would have to be carried out by conventional triple, quadruple effect equipment, or by vapor compression. To prevent scaling of the heat exchange surfaces pretreatment or cleaning methods would have to be employed, and concentrations would have to be kept below saturation.

For concentrations from 40% to dryness, evaporation would have to be carried out by special drying equipment such as submerged combustion evaporators which are non-scaling, flash dryers, or crystallizing evaporators. Because of their low thermal efficiency and mechanical complications, these units are generally very costly to operate.

Previous work on concentration by conventional evaporators has been done by Koenig and others (Koenig, '63). Koenig has presented his costs in Figure 1 of the reference cited, which is reproduced here for convenience (Figure 1). A utilization factor of 0.9 has been used, defined as the ratio of average production to the design capability.

In Figure 1 the units described have not been optimized for design and performance applicable to wastewater renovation. It should also be noted that economic factors today are different from the economic factors prevailing in 1962. However, the general approach is valid, and the work serves as a good background for accurate costing involving the actual engineering criteria and ground rules used in the current study.

A complete bibliography of the general background of disposal of brine effluents recently issued by the OSW (DePuy, '69) is included in the Bibliography of this report. See also the Appendix.

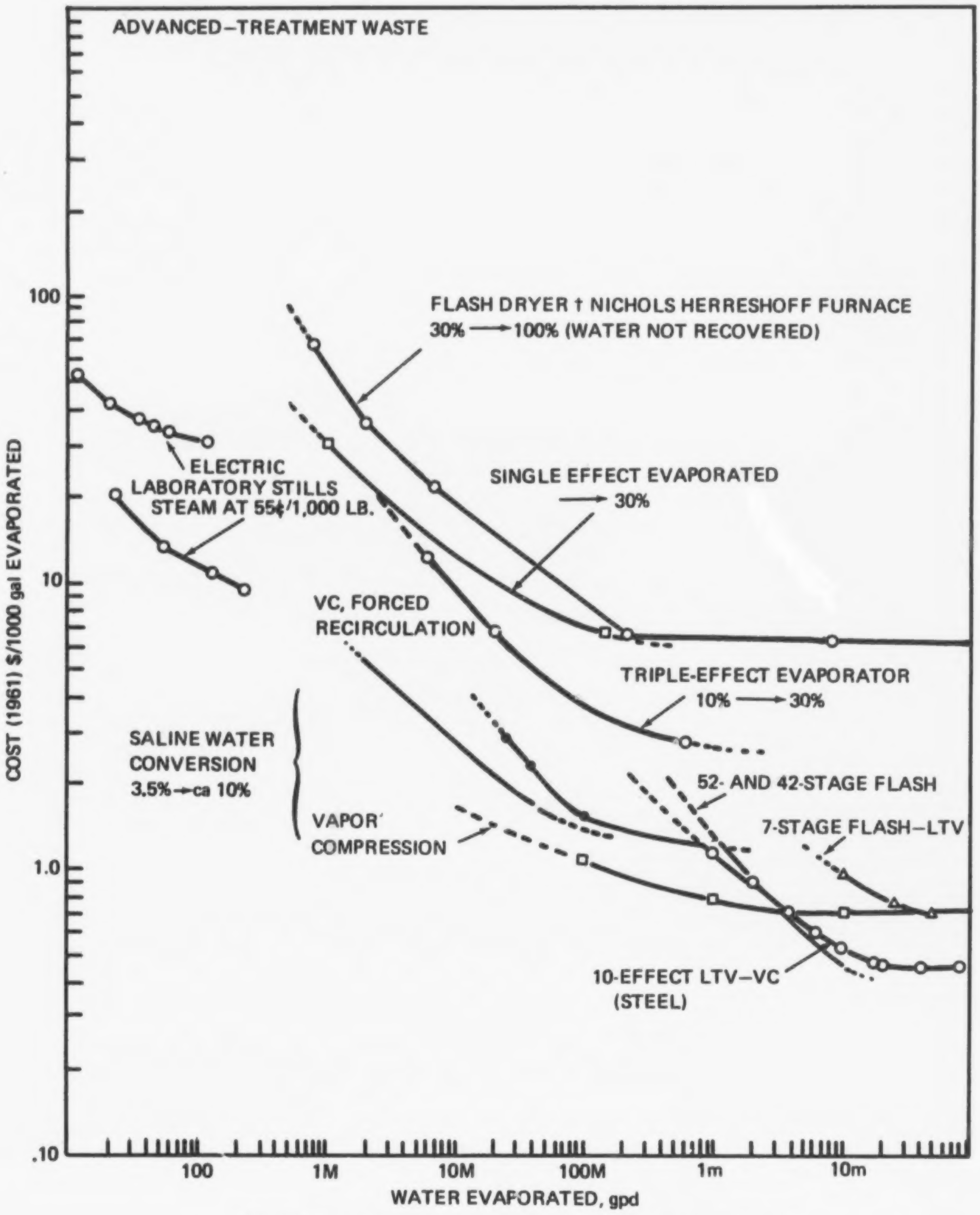

FIGURE 1. COST OF EVAPORATING WATER FROM SALINES

EVAPORATOR CONSTRUCTION MATERIAL: STAINLESS STEEL
OPERATION: 330 DAYS PER YEAR
ELECTRIC POWER: UNDER 100,000 kw, 7 MILLS/KWH
COOLING WATER: 40° F RISE
FUEL: 25¢/1,000,000 BTU
STEAM: 55¢/1,000 LB.
DIRECT LABOR: $2/HR.
GENERAL OVERHEAD, ADMINISTRATION AND PAYROLL EXTRAS:
45% OF DIRECT LABOR
MAINTENANCE MATERIALS, LABOR, AND OPERATING SUPPLIES:
2% OF TOTAL PLANT COST ANNUALLY
FIXED CHARGE RATE 7.4%

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DISPOSAL METHODS

The following paragraphs describe the important factors considered in the study of each of the processes. Rather than using a set of fixed economic conditions, as previous studies have done, the present study has developed, for each method of disposal, a set of parametric equations, which allow considerable variation in the economic parameters. Fixed charge rates over different lifetimes and for both public and private utilities, and electric power rates for large and small installations and for different parts of the country, have been considered. A range of steam costs applicable to all types of fuel has been developed, as well as capital costs for different materials of construction, and escalation factors for future installations. At the end of this report is a nomenclature, listing all of the symbols used in the mathematical developments.

DEEP WELL INJECTION

Abstract

The required pump pressure depends primarily upon the porosity and permeability of the injection strata and not upon the pipe size of the well. Therefore, the pumping cost and the capital cost are primarily dependent upon capacity and well depth, and not upon optimization parameters. Thus, well injection unit costs have been developed as functions of capacity and well depth, based upon practical considerations, rather than upon an optimization. Different assumed well-head pressures result in different costs. Unit costs for well injection have been calculated for a grid of economic factors, based upon fixed charge rates and electric energy costs. Necessary parametric equations have been developed.

Introduction

Injection disposal of waste brine involves comprehensive geologic investigation and field testing of the disposal zones and reservoir areas to determine that safe, effective underground disposal is possible, that the injected brine can be physically and chemically treated so as to be compatible with fluids in the reservoir, that there is no danger of pollution of or encroachment on underground fresh water supplies, that future natural resources will not be compromised by injection and that risks of unforeseen difficulties, such as earthquakes and eruptions are minimized.

Based upon the geological information available for the three sites of this study, injection wells cannot be used at all in the Denver area because of the earthquake hazard. At Tucson, Arizona, however, and also at El Paso, Texas, suitable injection strata are found at about 3500 feet depth. At both these locations, water is being taken from the reservoirs at a faster rate than recharge occurs. As a consequence, the water table is being lowered in both areas. Statewide water plans call for importation of fresh supplies before this decade ends, which will at least partially alleviate the shortage. It seems possible in the interim to conserve fresh water supplies by wastewater renovation techniques employing desalination and recharge. This move would tend to stabilize the water tables to some degree. Furthermore, if the waste brine from the desalination plant is injected during the interim into saline strata underlying the fresh water supplies, the result would be to displace the existing fresh water upward and to raise the water tables. This practice could be continued up until the time that sufficient imported fresh water supplies become available for complete recharge of the reservoirs. Then another method for disposal of the brine would have to be found.

Before injection into the strata, the waste brine should be treated physically and chemically to render it compatible with fluids in the reservoir. Typical treatments would involve desulfation of brines injected into formations containing Ba^{++} and Sr^{++}, deaeration of brines injected into formations containing Fe^{++}, and softening of brines injected into carbonate formations.

Conceptual Design of the Injection System

A drawing of the injection well is shown in Figure 2, for a 200,000 gpd well capacity. A 4-1/2" diameter injection tube is carried inside a 7" inside diameter casing long string to the 3500 foot level, and is sealed with well cement to form a pressurized annulus around the injection long string. Any change of pressure in the annulus is indicated by a pressure gauge, and warns against leakage. This serves to protect the intermediate fresh water bearing strata from contamination with brine. The well hole communicates directly with the injection long string and extends to a suitable depth beneath it to allow the required flow of brine.

The overall concept is that of a well field supplied by treated waste brine from an inland wastewater renovation facility. Figure 3 shows a typical arrangement. From the desalination facility, the waste brine flows first into a stabilizing pond. Here simple treatments, like aging and pH adjustments are employed to render the brine compatible with fluid already in the reservoir. Stabilized brine from the pond is transferred by means of a transfer pump to the injection well area, which, depending on the local geology may be either adjacent to or remote from the desalination facility. The brine is pressurized to the necessary well-head pressure by suitable injection pumps. Following these, guard filters are usually provided to remove suspended solids that otherwise could plug the permeable injection strata. From the guard filters, the brine flows into the distribution headers of the well field.

Deep Well Injection Pressure Relationships and Calculations

The operating well head pressure is the vector sum of the bottom hole injection pressure, the pressure due to the height of the injected brine column, and the pressure drop due to pipe friction. The bottom hole injection pressure is the sum of the reservoir pressure and the bottom hole driving pressure, or pressure drop due to formation friction. Generally, the operating well head pressure, or surface injection pressure, is limited to 0.5 psi per foot of depth to the injection zone.

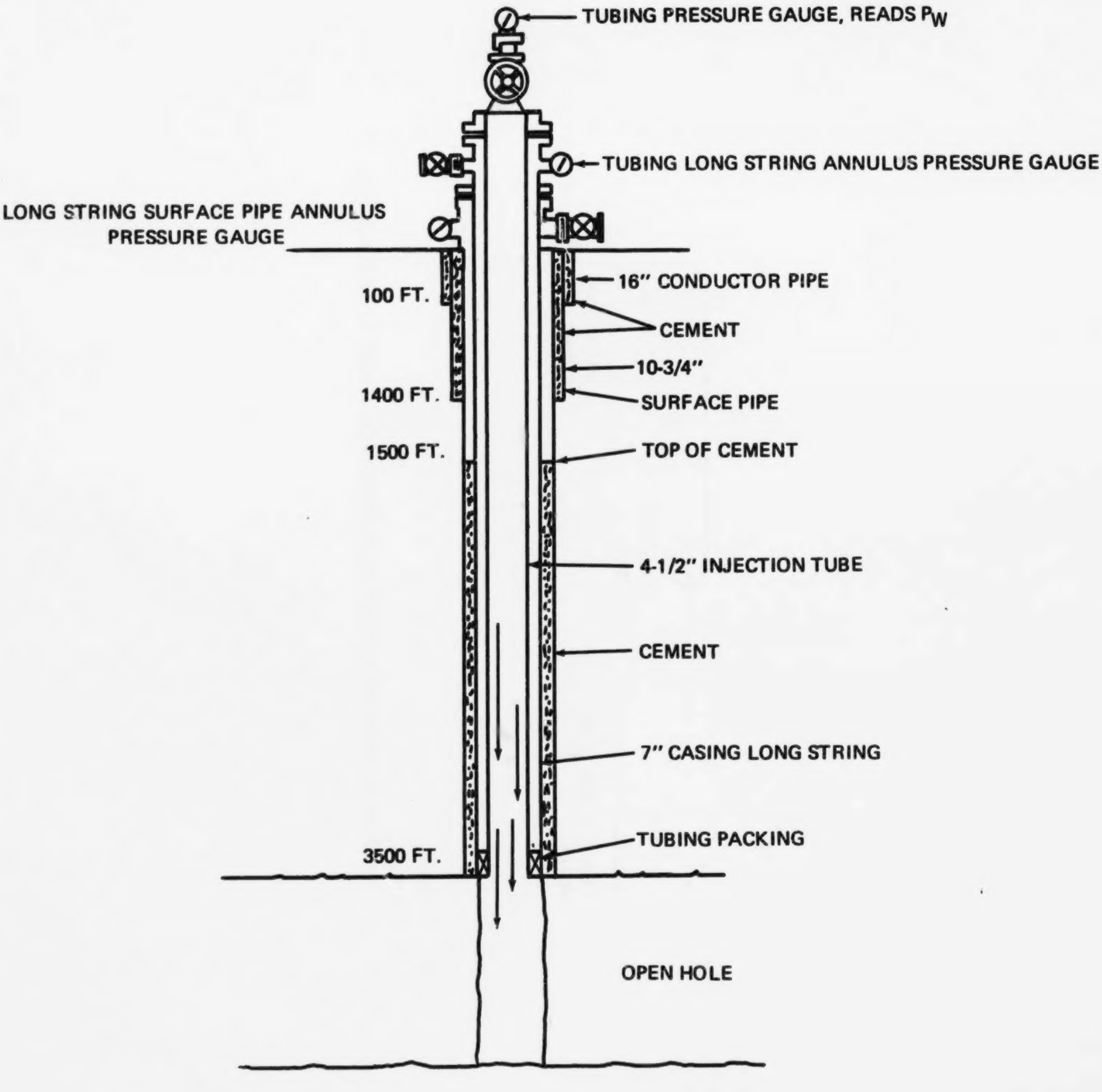


FIGURE 2. HUECO-BOLSON BASIN, DEEP WELL INJECTION DESIGN

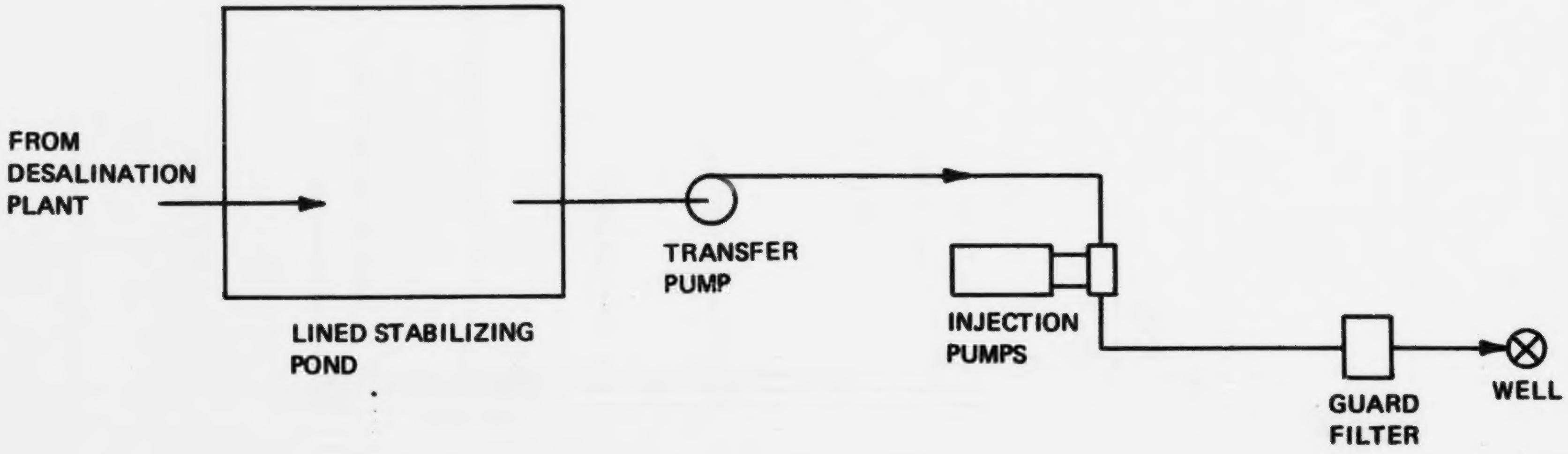


FIGURE 3. DEEP WELL DISPOSAL SURFACE EQUIPMENT FLOW DIAGRAM

Thus, for a 3500 foot well depth, the maximum well-head pressure is limited to 1750 psi. This limitation is based on the consideration that the bottom hole injection pressure plus the hydrostatic head of the injected fluid should not exceed the fracture pressure.

Both at Tucson and El Paso the permeability of the strata at 3500 feet is sufficient to allow introducing the brine at relatively low well-head pressures. At some places, in fact, gravity alone suffices for injection at this depth, so that a near zero well-head pressure can be employed. This condition cannot be expected to last over the lifetime of the well, however, and so in this study we have assumed that a back-pressure exists at the well bottom, which including the injection driving pressure is equivalent to a fluid in the stratum with a specific gravity of 1.116 and a head of 3500 feet.

Therefore, for injecting 700 mg/l brine with a specific gravity of 1.04 at a 3500 foot depth, the well-head pressure exclusive of pipe friction would be:

$$P_w^o = 3500\ (.4331)\ (1.116-1.04)\ \text{psig}$$

$$P_w^o = 169.75\ \text{psig}$$

Assuming a 4-1/2" O.D. injection tube for a 200,000 gpd well, and using the Hazen and Williams nomogram (p. 386 Ch.E. Handbook 3rd Ed.) the pipe friction is:

$$P_f = (2.2\ \text{ft}/100\ \text{ft})\ (35/2.303) = 33.4\ \text{psig}$$

Therefore the total well head pressure required is:

$$P_w' = P_w^o + P_f = 203.1\ \text{psig}$$

For a 100,000 gpd well with a 3-1/2" O.D. tubing the same well-head pressure is required.

Deep-Well Injection Cost

The cost of brine disposal by deep well injection is the sum of the following components: cost of wells and surface equipments; cost of well field; cost of operating, labor and maintenance and the cost of pipeline conveyance. However, the pipeline cost can be found in the pipeline conveyance cost section of this report, and is not included in the following well injection cost summaries.

Injection Wells and Surface Equipment Cost

For a 3500-foot deep well and a design capacity of 0.2 mgd at an operating pressure of 204 psig using a 4-1/2" O.D. injection string, the estimated cost is $66,900. An 0.1 mgd capacity well using a 3-1/2" O.D. injection string, is estimated to cost $63,380. (See Table 2.)

An additional cost is due to the pump and valves. With a pump efficiency of 0.85 and a motor efficiency of 0.93, this amounts to $8500Q, where Q is mgd well capacity. Since an 0.1 mgd well costs almost as much as an 0.2 mgd well, the unit well cost for the odd gallonage is almost twice that of even. Expressing the capacity of the entire injection field by Q leads to two equations depending on whether the field capacity in 0.1 mgd is odd or even.

Injection wells and surface equipment cost, exclusive of the well field distribution piping, is as follows:

(1a) $\$ \text{ Well Cost} = 343000Q$ (all wells are 0.2 mgd each)

(1b) $\$ \text{ Well Cost} = 343000Q + 29930$ (one well is 0.1 mgd)

General Equation for Well Field Costs (SW Res Inst. '66)

Refer to the appendix for layouts of different well fields from 0.1 mgd to 10 mgd capacity. The capital costs for well fields are given by two equations. These are based on multiples of the 0.2 mgd well, which is the maximum capacity for a single well in the well field:

(2a) For 0.1 to 0.9 mgd

$$\$ \text{ Well field piping cost} = (19303)\,(HxP)^{-1/2}\,(1.6558Q^2+5.06Q-0.708)$$

(2b) For 1.0 to 10 mgd

$$\$ \text{ Well field piping cost} = (19303)(HxP)^{-1/2}(0.382Q^2+8.9612Q-3.4773)$$

General Equation for Operating Cost, excluding Capital Recovery

$$\$/\text{Kgal.} = .00777/Q + .00187Z + .0359 - .00048Q$$

General Equation for the Total Brine Disposal Cost

$$\$/\text{Kgal.} = 2.74(10)^{-6}(FCR)\ \Sigma\ \$\text{Capital Cost}/Q + .00777/Q + .00187Z + .0359 - .00048Q$$

Conclusions and Summary

The following graph (Figure 4) shows total brine disposal costs calculated from this equation for a 3500 foot deep well with operating pressure of 204 psig at various well field sizes and fixed charge rates.

NOTE: Capital costs are determined from equations (1a,b) and 2a,b).

TABLE 2

ESTIMATED COST OF INJECTION WELL CONSTRUCTION

1. Drilling to 3100 feet	$13,400
2. Contractor day work	9,000
3. Tubular goods	
a. 100 ft 16-inch conduction pipe	1,500
b. 1400 ft. 10-3/4 inch surface pipe	5,600
c. 3500 ft 7-inch long string	8,420
d. 3500 ft 4-1/2 inch injection pipe with PVC lining	7,580
e. Well head and packer	4,000
4. Well services	
a. Cementing and associated equipment	4,000
b. Wire line logging and perforating	2,500
c. Formation testing, core analysis and testing	2,500
d. Acidizing, mud and chemicals	3,400
5. Miscellaneous	
a. Tubing rentals	1,000
b. Engineering and geological service	1,000
c. Location work, trucking, travel, etc.	3,000
Total Cost for one 0.2 mgd well	$66,900
Similarly, for one mgd well, the cost for a 3500 ft and 3-1/2 in./O.D. injection pipe with PVC lining is $4,060 - Thus, Total Cost for one 0.1 mgd well is	$63,380

Reference (Dow Chem. '69)

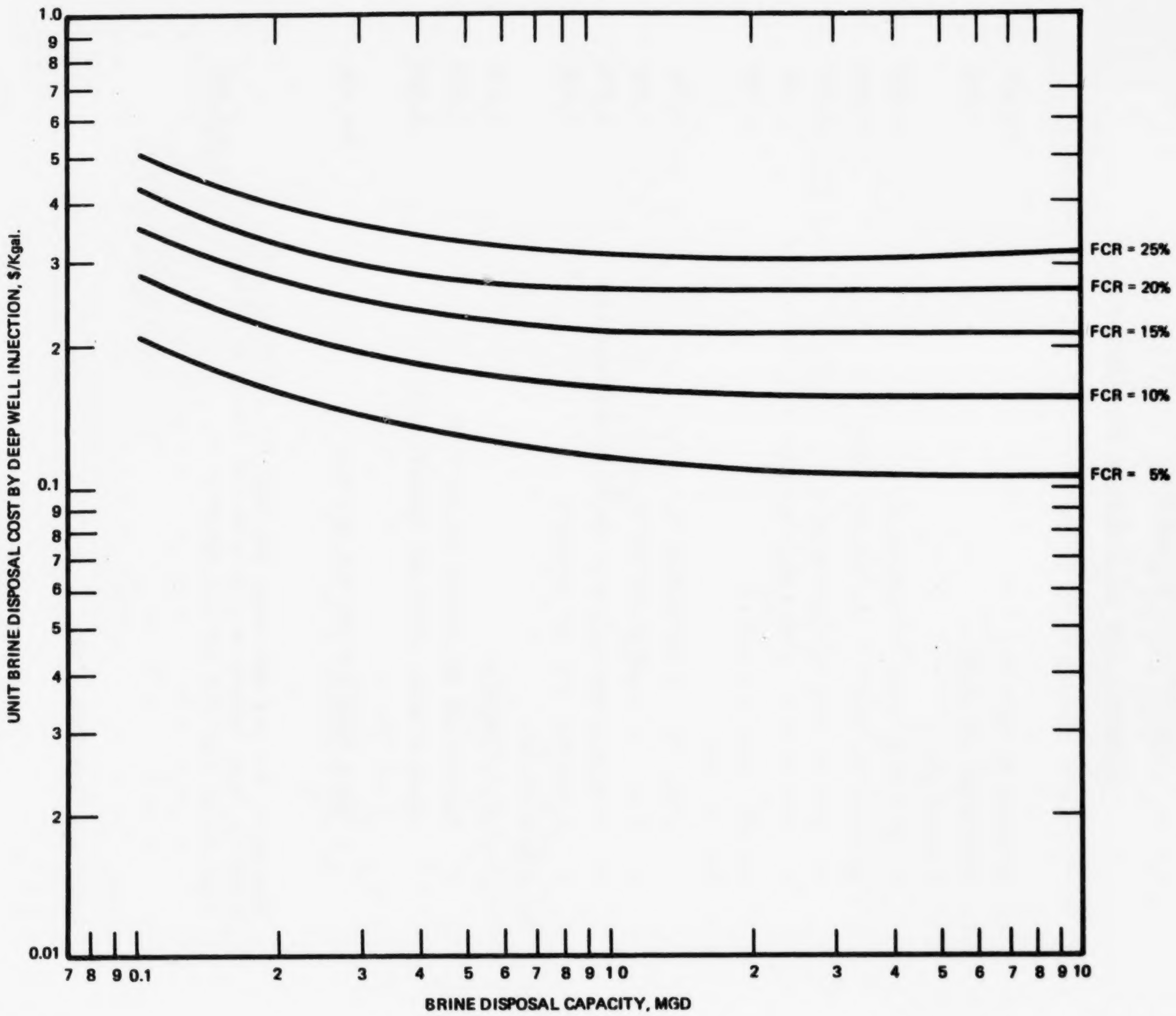


FIGURE 4. DEEP WELL INJECTION

SOLAR EVAPORATION PONDS

Introduction

Solar ponds are feasible only where annual evaporation exceeds the annual rainfall by over 20", as it does in the far West. Practical considerations govern the size and costs of the solar ponds. The costs are functions of the concentrating factor, capacity and net annual evaporation rate. Three sets of costs have been developed: (1) a cost for concentration to near saturation for ultimate disposal; (2) a cost for evaporation to near dryness for abandonment at the evaporation site, and (3) a cost for preconcentrating to saturation plus pipelining the residue away to the abandonment site, for which a computer program has been developed.

In many arid areas it is possible to evaporate brines to dryness in properly lined solar evaporation ponds. The general cost equations presented in this section have been developed upon the basic assumption that the evaporation pond would be located at the desalination plant without a pipeline between the pond and the plant.

Solar evaporation pond costs include the original cost of the land, the cost of stripping the land and providing dike material and fill and the cost of lining, which is the major cost in this process. Each of these cost components is described briefly below.

Calculations of Size and Cost

Total Area of Land: In order to accommodate the width of the dike, the total area of land required is slightly greater (about 3% to 5%) than the required waste brine surface area. Land cost varies from \$50 to \$500 per acre. A \$250 per acre cost has been used as the standard land cost in this study.

The waste brine surface area is dependent upon the annual net evaporation rate and the waste brine input rate:

$$AE = \frac{1.344 \times 10^4 Q}{ER} \quad \text{in acres.}$$

A safety factor of 24% is added to allow for the reduced evaporability of salt water and the area occupied by dikes, therefore, the total land requirement becomes:

$$T_a = 1.24 \times AE, \text{ acres}$$

$$\text{Cost of land is CLND} \times T_a$$

Use of Ta area will result in a saturated brine over the lifetime of the wastewater renovation facility, and a concentration to dryness after abandonment.

Dike Material: The pond is designed to accommodate the waste brine precipitate as a solid for a period of 30 years. A 3-1/2' basic pond depth is assumed to allow for heavy rainfall years. It is assumed that this pond would be square and the dike would have a four-foot crest with a 2:1 slope on the toe and a 3:1 slope on the heel.

The cost of the dike is assumed to be $1.00 per cubic yard including equipment, labor and material. The volume of dike material depends upon the total dike height, the total length of the dike and the thickness of the dike material.

Total dike height depends upon summing the following:

Basic pond depth: 3.5 (ft/yr)

Precipitate depth (ft):

$$P_d = P_t \left(\frac{\text{ft precip}}{\text{ft evap}}\right) \times \frac{ER}{12} \left(\frac{\text{ft evap}}{\text{yr}}\right) \times \text{Pond Life (years)}$$

A 3.0 ft surge capacity depth, freeboard which would be maintained to accommodate the summer to winter evaporation rate variation.

Freeboard for waves. The wave height can be found by the use of the Stevenson's formula.

$$HW = 1.5 \sqrt{F} - \sqrt[4]{F} + 2.5$$

However, to simplify the calculation, we assumed an average value of 3.0 ft. Here F, the evaporation pond "Fetch" is taken equal to the pond length.

Liner Cover. A one-foot soil cover would be placed on the liner for protection from damage.

Hence: Summing these for a 30 year pond life:

$$HD = \frac{30}{12} \times ER \times P_t + 10.5$$

$$HD = 10.5 + ER\,(2.5\,P_t)$$

But, for brine concentration of 7000 mg/1, $ER\,(2.5\,P_t) \cong 0.5'$ (OSW '66)

Thus, $HD = 11.0$ ft.

Total length of the dike is $L = 4 \times \sqrt{4840 \times AE}$, yards

Dike Volume

$$V_D = (L \times HD/9)\,(4 + 2.5 \times HD),\ yd^3$$

Dike cost will be: $\$1.00 \times V_D$

Lining: The area of liner is the total length of the dike multiplied by the total pond height plus the total base area. In order to simplify the calculation, the lining area is assumed to be approximately equal to the total land area:

$$ALNR = 43{,}560 \times T_a\ ft^2$$

Cost of lining is as follows:

30 mils (PVC) lining is $\$0.132/ft^2$ based on 60¢/lb

Liner cost = CLNR x ALNR

Volume of Fill to Cover Liner: Cost of one foot of soil to cover liner is assumed to be \$0.40 per cubic yard, including equipment, labor and material. Total volume of fill will then be:

$$V_f = \frac{ALNR}{27},\ yd^3$$

Cost of liner cover is: $\$0.40 \times V_f$

Stripping of Land: Cost of stripping land is assumed to be \$100 per acre. The cost will then be $\$100 \times T_a$.

The total cost for a solar evaporation pond is:

CE = Land cost + liner cost + stripping land cost + liner cover cost + dike cost

$$CE = CLND \times T_a + CLNR \times ALNR + 100 \times T_a + 0.4 \times V_f + 1.00 \times V_D$$

Conclusions and Summary

Substituting all intermediate relations for CE, the unit cost \$/Kgal. of disposal can be expressed in the form:

$$CE_u = B_4 + B_5 \times \frac{1}{\sqrt{Q}}$$

where

$$B_4 = 2.74 \times 10^{-6} \times \frac{FCR}{ER} (1.667 \times 10^{4} \times CLND + 7.26 \times 10^{8} \times CLNR + 1.242 \times 10^{7})$$

$$B_5 = 2.74 \times 10^{-6} \times \frac{FCR}{\sqrt{ER}} (1.242 \times 10^{6})$$

Similarly for unlined ponds the unit disposal cost in \$/Kgal. of brine is:

$$2.74\ (10)^{-6} \times FCR \left[\frac{1}{ER} (1.667\ (10)^{4} \times CLND + 1.667\ (10)^{6}) + \frac{1}{\sqrt{Q \cdot ER}} (1.039\ (10)^{6}) \right]$$

The following graph (Figure 5) shows costs of disposal by solar evaporation as a function of fixed charge rate and capacity in mgd at a site where the net annual evaporation rate is 90". Costs for other conditions can be obtained by use of the foregoing equation.

CONCENTRATION TO SATURATION AND PIPELINING

Introduction

In considering ultimate waste disposal costs, not only must each different method be evaluated, but combinations of methods must be studied. The scope of this report specifically requires investigation of the most promising combinations of methods.

The purpose of this section is to investigate a solar pond in combination with a pipeline. The objective is to reduce the volume to be disposed of by means of the solar pond and to pipeline the concentrated brine to a site where the ultimate disposal will take place (a salt lake or injection well, for example).

The cost to be evaluated and minimized is the total cost in \$/Kgal. for the disposal of the brine. This cost is the sum of the costs of concentrating the brine in a solar pond (to reduce the volume) and conveyance of the concentrated brine to the battery limits shown in the figure below.

Evaluation

Q mgd per day of waste, with a TDS concentration of S, are fed into a solar pond from the treatment facility as shown in Figure 6. In the pond the volume is reduced to Q_p, and the concentration is increased to S_p. From here Q_p is then transported to the battery limits via a pipeline.

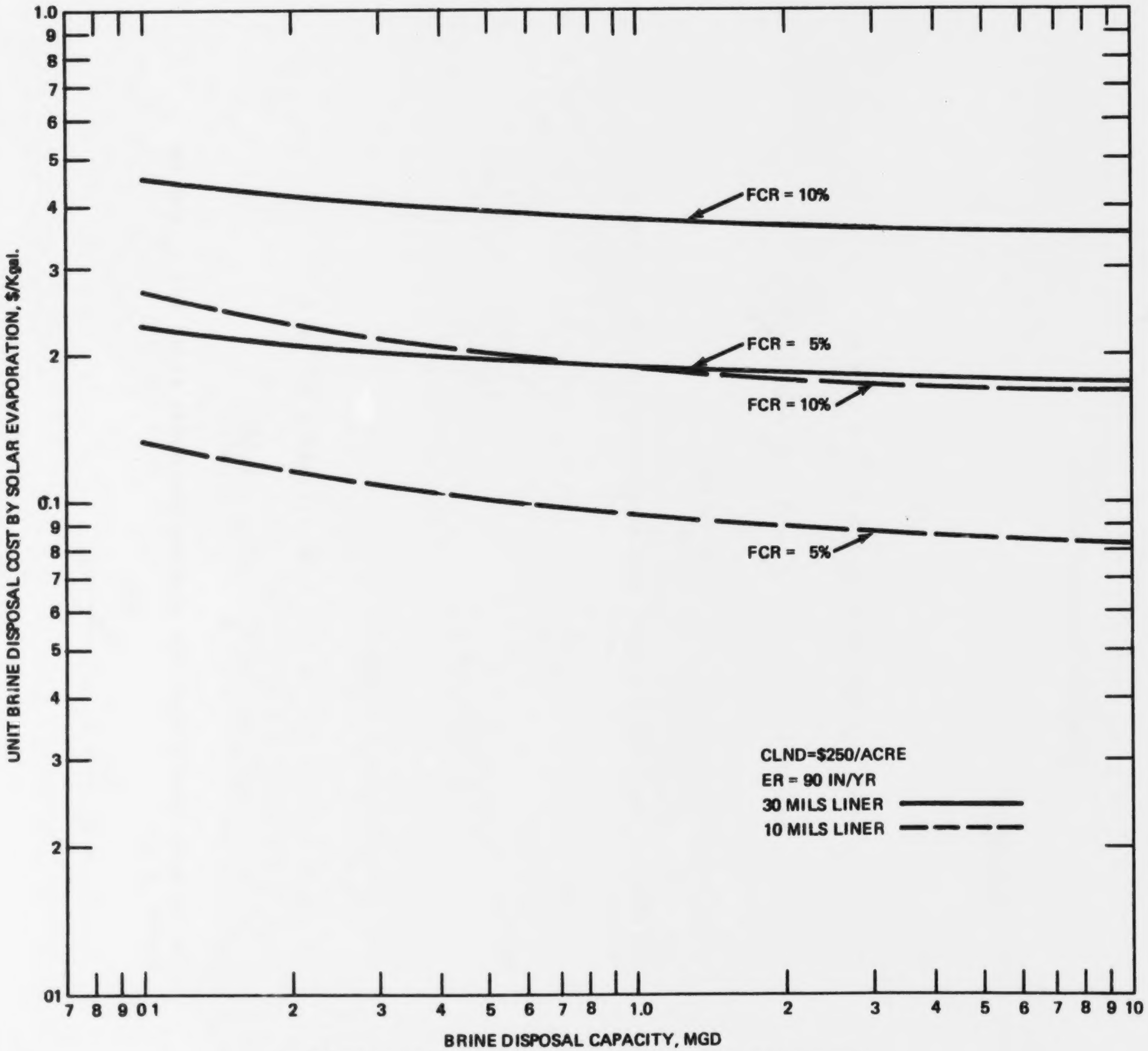


FIGURE 5. SOLAR EVAPORATION OF BRINE TO EVENTUAL DRYNESS

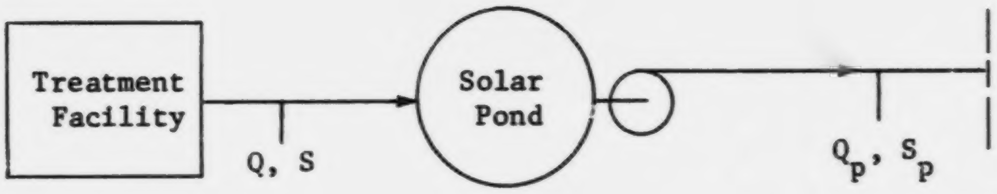


FIGURE 6 DISPOSAL SCHEME

Mathematical Formulation:

Let: $\frac{S_p}{S} = S_r$ = Concentration ratio

Therefore, flow in pipeline = $Q_p = Q \times \frac{S}{S_p} = \frac{1}{S_r} \times Q$

amount evaporated = $Q_e = Q - Q_p = (1 - \frac{1}{S_r})\ Q$

Total cost of disposal: $CT_u = CP_u + CE_u$

The partial unit cost \$/Kgal. for pipelining ℓ miles can be shown to be:

$$CP_u = B_3\ Q_p^{\ 0.615} \times \frac{10^{-3}}{Q}$$

Here

$$B_3 = \frac{A_1 x A_3 + A_2}{(A_3)^{5/6.29}} \times 10 \times \ell$$

$$A_1 = 1.447 \times A \times FCR'$$

$$A_2 = 5.628 \times 10^{-4} \times FCR' \times B* + 2.683 \times 10^{-3} \times Z$$

$$A_3 = (1.508 \times 10^{-3} \times \frac{B*}{A} + 0.719 \times \frac{Z\ (10)^{-2}}{AxFCR'})$$

The partial cost \$/Kgal for evaporation ponds sized for Q_e can be shown to be

$$CE_u = B_4 \frac{Q_e}{Q} + B_5 \times (\frac{Q_e^{\ 1/2}}{Q})$$

Replacing Q_p and Q_e, by Q and S, we get the total cost of disposal at ℓ miles, \$/Kgal.

$$CT_u = \frac{.001 \times B_3}{Q^{.385} \, S_r^{\,.615}} + B_4 \left(1 - \frac{1}{S}\right) + B_5 \sqrt{\frac{1 - \frac{1}{S_r}}{Q}} \quad \$/\text{Kgal.}$$

Conclusions and Summary

The computer program in the Appendix reveals that:

The cost of disposal for the combined solar pond and pipeline method can be compared with others and the most economical disposal method can be chosen. At different sites (different economic conditions), different methods turn out to be the most economical. Even at the same site the method chosen depends on capacity. For fixed economic conditions, as the capacity increases preconcentration costs must decrease in order to get a cheaper cost for the combined method than for pipelining alone. Therefore, there exists a capacity above which pipelining alone with no preconcentrating should be used. Also at some sites evaporation may always be cheaper than pipelining. In such cases, abandonment in lined ponds at the evaporation site should be considered.

In this study we assumed that the concentrated brine was dumped on useless land. Obviously any added cost for dumping the brine will affect the optimization.

Costs of brine disposal \$/Kgal. at 50 miles for a 90" ER, and 12 mills power cost are shown on the following graph (Figure 7). Other costs can be obtained by direct substitution into the above equation, preferably using a computer.

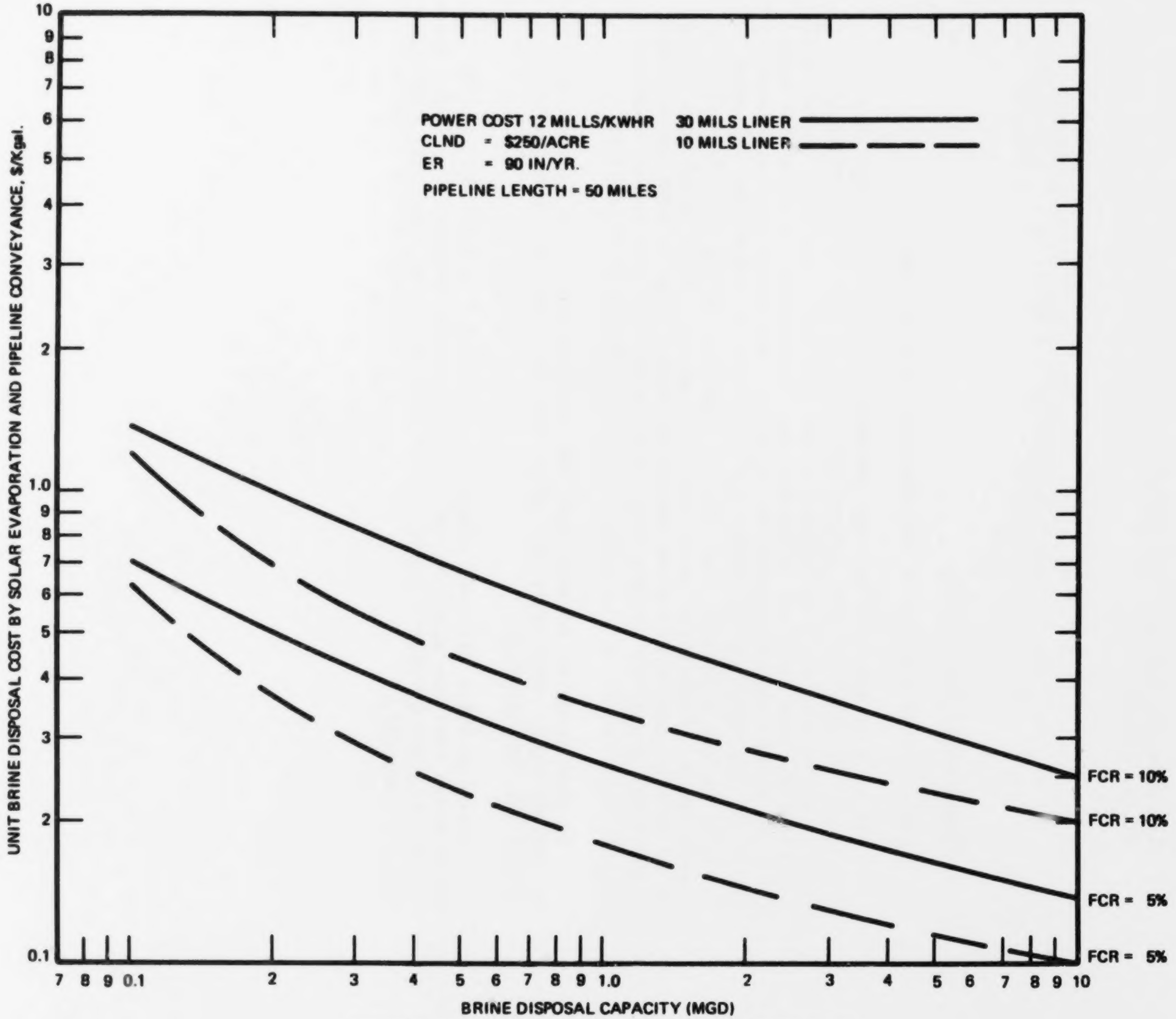


FIGURE 7. SOLAR EVAPORATION AND PIPELINE CONVEYANCE

MULTISTAGE FLASH EVAPORATORS

Abstract

Optimum geometries for multistage flash evaporators have been developed based on the minimum capital investment necessary for the desired production rate at the optimum performance ratio. A nomogram has been developed which shows the optimum number of stages and performance ratio as a function of the plant size, the annual fixed charge rate, the electrical power cost, and the steam cost. Unit costs and capital costs for preconcentration by evaporation for ultimate disposal have been shown on a second nomogram based upon annual fixed charge rates, unit steam costs, plant size, mgd evaporated, and optimum performance ratios. Costs are expressed per 1,000 gallons per day evaporated.

Introduction

Recent work in the art of desalination has progressed to the point where an accurate economic assessment of the costs incurred for producing fresh water from brackish water can be made. Conversely, the cost of concentrating brackish water can be accurately determined. Consider a brackish stream entering a black box, with two exit streams. One effluent stream is concentrated brine, and the other is pure water. The cost of operating the black box (evaporator) includes three basic costs:

Capital cost of equipment
Cost of fuel
Other operating costs

These three cost parameters determine the cost of separating the influent brine stream into the two effluent streams. It now becomes a matter of accounting procedure to determine the cost of each effluent stream. For example, assume the concentrated brine is the primary product, and the total cost of separation is \$0.50/Kgal. If the pure water produced has zero value and is pumped to waste via a river, then the entire cost of separation must be borne by the concentrated brine. If, however, it is found that the pure water can be sold for agricultural purposes at \$0.20 per thousand gallons, then the cost of concentration is \$0.50-\$0.20=\$0.30 per thousand gallons. Obviously, if the disposal costs are greater than \$0.30 per thousand gallons, evaporation may be a usable disposal technique.

Purpose and Concept

Multistage flash evaporation is the most economical fully commercial method available for concentrating inorganic salts up to 10% tds. In this process the cost per thousand gallons of water evaporated is of interest, and the blowdown is to be disposed of by other means. The minimum cost per thousand gallons of water evaporated can be obtained by finding the performance ratio (defined as the pounds of water evaporated per 1,000 Btu's supplied to the brine heater) at which the total operating costs are minimum for a given set of operating and economic parameters. The selection of a suitable set of operating parameters for concentrating waste brines from municipal wastewater renovation plants is the first task described in this section. The second task is the selection of a general grid of economic parameters to allow accurate costing of evaporation for any site selected within the United States. The next task is to establish optimum performance ratios for the entire grid of economic conditions. The final task is the conceptual design of a multistage flash evaporator for the specific purpose of concentrating waste brines from municipal wastewater renovation plants.

Operating Parameters

Where scaling of the brine heater is not a problem, previous studies (Fluor, '59; Bechtel, '66) have demonstrated for single purpose plants that:

> The practical blowdown concentration for seawater conversion evaporators is between 6.0 and 7.0 percent dissolved solids, with the total optimization being rather insensitive to this parameter. For waste brine evaporators it is expected to be about 10%.
>
> The optimum brine-heater outlet temperature should be about 350°F.
>
> The evaporation plant pumps should be turbine-driven, and the turbine exhaust steam should be used for the brine heater. This results in the lowest power generation costs.
>
> Techniques have been developed for optimizing the brine heater and the brine blowdown temperatures. In no event should the brine blowdown temperature be allowed to fall below 85°F, because of difficulties in removing non-condensables at such low temperatures.
>
> In cylindrical evaporator vessels, overflow type flashing weirs, as described by Mulford (1965), should be run longitudinally (e.g., parallel to the tubes) between sloping floor sections in order to provide the maximum economy of construction and still limit the maximum brine overflow rate to less than 800,000 pounds per hour per foot of weir. In the coldest stages, the weirs should be longer, so that a brine rate of not over 400,000 pounds per hour is achieved.

In the current study of concentrating the brine from a municipal wastewater renovation facility it has been assumed that the preceding process is electrodialysis which concentrates to 7,000 mg/l for feed to multistage flash evaporation. It has been further assumed that a post treatment has been used in the electrodialysis plant, so that neutralization to 7.0 pH of the 7,000 mg/l influent brine to the multistage flash evaporators will allow scale-free operation of the brine heater at 350°F with blowdown concentrations up to 10% at a temperature of 100°F.

Using the foregoing assumptions and operating parameters, the flowing quantities for four sizes of ultimate disposal plants are:

Description	Small	Intermediate		Large
AWT WASTE BRINE at 7,000 mg/l, MGD	0.1	1.0	2.5	10.0
BLOWDOWN at 100,000 mg/l for ULTIMATE DISPOSAL, MGD	0.007	.07	0.175	0.7

Economic Parameters

In an evaporator plant, there are four economic parameters to be considered:

Capital Cost, for which the total investment cost is used;

Fixed Charge Rate (FCR), or principally the capital recovery factor;

Cost of Heat, or steam cost in \$/mbtu at the brine heater;

Pumping Cost, for shaft work in \$/Kgal. of production.

Capital Cost: The total investment cost, including engineering fees and profit, site development and land, must first be itemized to show costs that depend upon the performance ratio (lbs water/1000 Btu at the brine heater), and costs that are independent of performance ratio because they depend upon capacity only. The accuracy of this itemization determines the accuracy, and in fact, the validity of the entire optimization scheme. For the operating parameters previously described, the total erected cost of the evaporator vessels depends on the plant capacity and the performance ratio. These costs can be divided into two parts: those proportional to condenser area and those proportional to weir length. The cost of condenser bundles, demisters, and product water trays is given (using Burns and Roe costs) by the equation:

$$\$/\text{ft}^2 \text{ condenser area} = 7.5/(\text{MGD})^{0.22}$$

The cost of stage partitions, floors, and flashing weirs can be expressed as a function of weir length. Development of these costs is illustrated for a 2.5 MGD plant in Table 3. Weir length is equal to four times plant capacity in MGD and is 10 feet for a 2.5 MGD plant. This results in a brine rate of less than 400,000 lb/(ft)(hr) which prices all plants conservatively.

The two components of erected cost of evaporator vessels are shown below for various plant sizes:

Plant capacity (MGD)	0.1	1.0	2.5	10.0
Evaporator cost, $\$/\text{ft}^2$ of condenser	12.50	7.75	6.25	4.50
Stage cost, \$/ft of weir	1400	868	700	504

The above costs are based on the 2.5 MGD plant size, and extrapolations are made to the other sizes using a Chilton factor of 0.78.

TABLE 3

TOTAL COST PER STAGE FOR A 2.5 MGD WASTEWATER DESALTING PLANT

Maximum Brine Temperature 350° F

COST PER STAGE:

	Material	Length ft	Width ft	Area ft^2	Gauge in.	Corrosion Allowance in.	Total in.	Unit Wt. lbs/ft^2	Total Wt. lbs	Fabricated Cost		Erection Cost		Total Cost $
										Unit $/lb	Total $	Unit $/lb	Total $	
Stage Separator Plate Top	C. Steel			63.84	1/4	3/16	7/16	17.85	1139	.90	1,015.10	.125	142.38	1,167.50
Stage Sep. Pl. Below Demisters	C. Steel	10	5	50	1/4	3/8	5/8	40.80	2040	.75	1,530.00	.125	255	1,785.00
Adjustable Flow Plate														100.00
Weir Plate	C. Steel	10	1	10	1/4	3/8	5/8	25.50	255	.75	191.25	.125	28.13	219.40
Splash Plate	C. Steel	10	3	30	1/4	3/8	5/8	25.50	765	.75	573.75	.125	95.63	669.40
Product Tray Ends	CuNi	60"	9"	1080 in^2			0.0508	.323 $\frac{lb}{in^3}$	17.44	1.60	27.90	.10	1.74	29.65
Sloping Floor	C. Steel	8-1/4	10	82.5	1/4	3/8	5/8	40.80	3350	.75	2519.00	.125	510.00	3,029.05
Total Cost														7,000.00
Cost per foot of weir														$ 700.00

Fixed Charge Rate: The fixed charge rate is the summation of the annual capital recovery factor, the annual insurance charges, and income and property taxes, all expressed as percentages of the total investment over the lifetime of the plant. A fictitious fixed charge rate based upon a uniform series must be developed if plant components are added or replaced during the plant lifetime, or if the actual components of the fixed charge rate are themselves non-uniform series. The fixed charge rate used for the optimizations herein included has a constant value over the plant lifetime, and this fact must be borne in mind when selecting this factor for an optimum plant at a given site. Values of .05 to .25 have been used in this study.

Cost of Heat: For the purposes of this study, all costs for generating saturated steam at 357°F, including the annual capital and operating costs of the boiler, and allowing for normal steam losses, must be summed and expressed as $/mbtu delivered across the brine heat exchanger. If the steam power plant is dual-purpose the cost of the steam will have to be decided arbitrarily, based upon local conditions; however, if it is single-purpose, turbine drives can be used for the water pumps, with their exhausts manifolded into the brine heater. All the generated heat that is not expended by pumping winds up in the brine heater. A typical scheme would employ a boiler generating 452°F steam at 245 psia pressure for supplying back pressure turbines. These turbines would exhaust steam to the brine heater of the water plant at 357°F and 147 psia. The boiler would be priced at about $1800/mbtu/hr at an applicable annual fixed charge rate. Fuel costs would be adjusted for an 84% stack efficiency with about a 3% allowance for loss of steam in ejectors and leaks, plus operating and maintenance labor on the boiler plant. Steam pricing must be done as carefully as possible, taking all local conditions into consideration, since this is the prime operating cost variable upon which the optimization depends. Selecting an incorrect cost for steam will result in an incorrect cost for water production and an off-optimum plant.

Pumping Cost: This cost has virtually no effect upon the optimization, inasmuch as varying the performance ratio of the plant causes an insignificant variation of the pumping head and no variation at all in the amount of water pumped. Therefore, it can be calculated as a fixed cost, dependent only upon the unit power charges in mills/kwhr and added to the variable water cost. For various power rates, the pumping cost per 1000 gallons of product water is as follows within the battery limits of the water plant:

Power Cost, mills/kwhr	5	7.5	10	12.5	15
Pumping Cost, $/Kgal. H_2O	.046	.0772	.092	0.115	.139

For turbine drives, the pumping cost will average about $.05/Kgal. per dollar cost of prime steam per million Btu's. Thus, if the steam cost used in the optimization is $0.50/mbtu, then the pumping cost for the optimized plant will be $0.025/Kgal. of product water.

Optimum Performance Ratios

In this study, performance ratios and stages have been optimized over a complete economic grid of capital costs, fixed charge rates, unit steam costs and plant capacities from 0.1 to 10 mgd. Refer to the Appendix for the mathematical development.

The optimization determines the cost of the water for the optimum evaporator package, including erection, for a given cost of steam. The cost of power, chemicals, maintenance, supplies and operating labor must be added, together with capital costs for product and recycle pumps and all other costs that depend on size and fixed charge rate. However, the costs of power, chemicals, and expendable supplies are independent of fixed charge rate. Annual maintenance and operating labor are independent both of plant size and fixed charge rates. These costs are included in the fixed cost for the water plant, except for operating labor, which we have not assessed because it will depend on local administrative procedure and cost sharing with the much larger wastewater renovation plant. Operating labor would have to be added as a separate item in any event.

Base Capital Costs for a 2.5 mgd plant:

Recycle pump	$ 30,000
Cooling water pump	20,000
Makeup	8,200
Blowdown and product pumps	8,000
Deaerator	51,000
Steam jet air ejector	9,220
Electric	95,800
Instrumentation	121,600
Piping	140,000
Facility cooling water intake and outlet piping	24,000
Chlorination	1,500
Site development	8,000
Buildings	40,000
Service water and sanitary	7,000
	$564,320
Total Markup including shipping and insurance, Engineering and Construction and Escalation @ 30% of Base Capital Cost.	169,300
Base Capital Cost	$733,620

Base capital cost for 0.1, 1.0 and 10 mgd can be calculated by simply multiplying a cost factor from the following tables by the base capital cost for 2.5 mgd or $734,000. Base water cost $/Kgal. is given directly by the table as a function of fixed charge rate and size.

mgd	0.1	1.0	2.5	10
Capital cost factor	.08	.495	1.0	2.92
FCR 5%	.0894	.0554	.0447	.0327
10%	.179	.111	.0894	.0654
15%	.268	.166	.134	.098
20%	.357	.221	.1785	.131
25%	.446	.276	.223	.163

Fixed Costs:

Chemicals, $/Kgal.	$0.016	(cost of neutralization to pH 7.0)
Maintenance and supply	0.004	
Total Fixed Cost	$0.02	$/Kgal.

Use of Nomograms for Optimum Performance Ratio and Water Costs

The economic conditions for waste brine disposal by evaporation are most simply shown by nomograms. These alignment charts are arranged for pivoting about a center key line, and cover a larger range of variables than required by the ground rules of this study.

Nomogram No. 1 allows the optimum performance ratio and the optimum number of stages to be determined from the unit cost of steam, the cost of condenser area and the annual fixed charge rate. It solves the basic equation for optimum performance ratio:

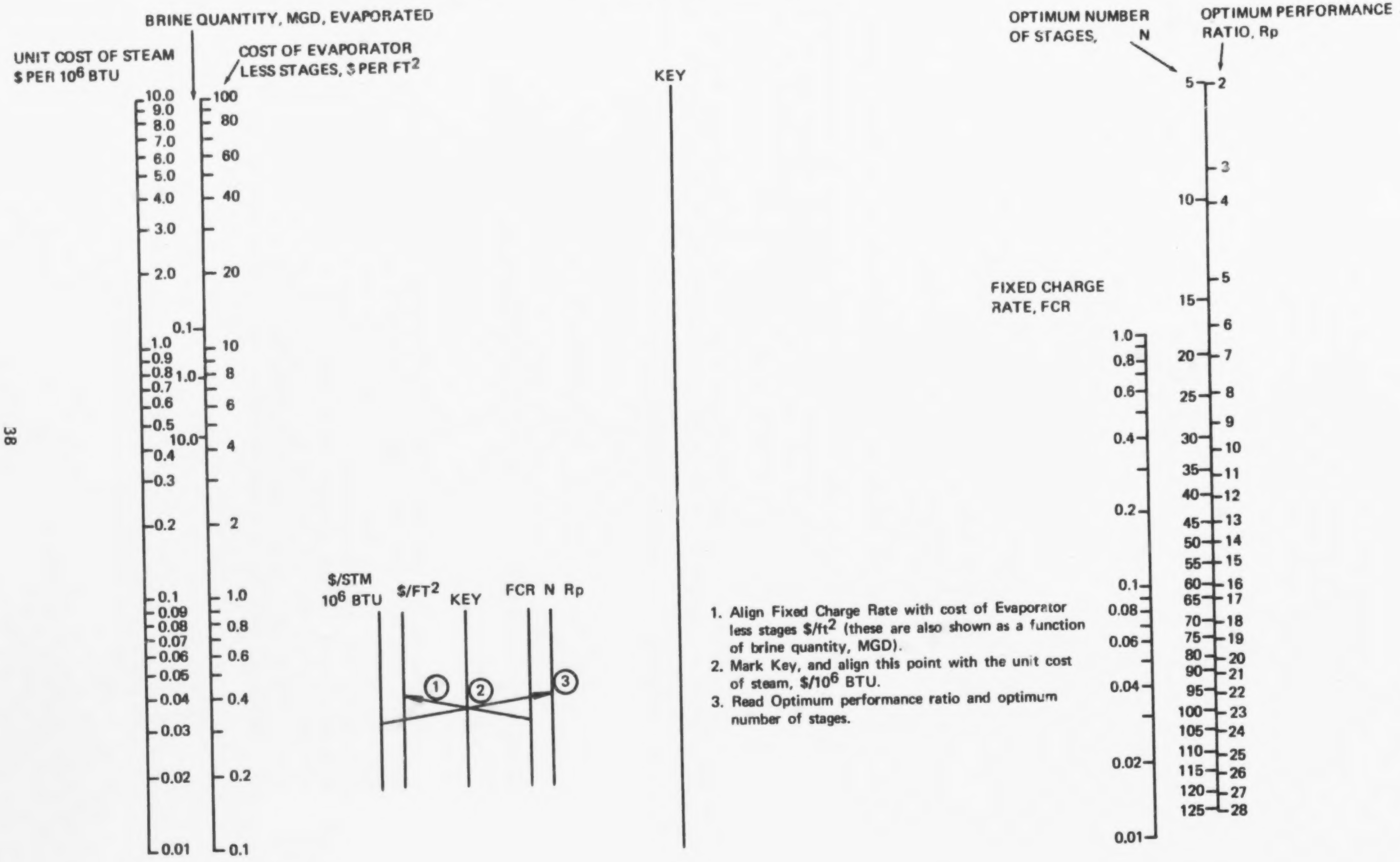
$$\frac{Rp^2 + Rp}{(1-.023Rp)^2} + 0.551(Rp)^2 = 846(\$Stm/\$A)(1/FCR)$$

along with the basic equation for the optimum number of stages:

$$N = 5.25Rp - 23.84$$

To use, align the applicable fixed charge rate (FCR) with the unit area cost ($\$A/ft^2$) for the evaporator condenser and shell. (These costs are also shown as a function of plant size, mgd, in which case the unit area costs can be disregarded, with the plant size being directly aligned with the fixed charge rate.) Mark the point so ascertained on the key, and then pivot on this point, so as to align the appropriate unit steam cost ($Stm/mbtu). Both the optimum performance ratio, Rp, and the number of stages, N, can now be read directly off of the nomogram.

OPTIMUM PERFORMANCE RATIO VS. BRINE QUANTITY, UNIT COST OF EVAPORATOR LESS STAGES, FCR AND STEAM COST, NOMOGRAM 1



1. Align Fixed Charge Rate with cost of Evaporator less stages $/ft² (these are also shown as a function of brine quantity, MGD).
2. Mark Key, and align this point with the unit cost of steam, $/10⁶ BTU.
3. Read Optimum performance ratio and optimum number of stages.

Nomogram No. 2 allows determination of the partial unit cost of water and the partial capital cost of the plant from the performance ratio and the cost of steam previously used, according to the equation:

$$\$H_2O/Kgal. = 16.6(\$Stm/mbtu)/Rp_{(opt.)}$$

First, align the optimum performance ratio from Nomogram No. 2 with the appropriate steam cost. Read off the corresponding partial cost of water, $/Kgal. (Fixed costs, base capital costs, power costs and operating labor must be added.)

Second, align this cost of water with the applicable fixed charge rate and mark the key line point.

Finally, align the key point with the appropriate plant size and read from the nomogram the partial capital cost of the evaporator and stages. (Fixed capital costs and markup must be added to get total investment costs.)

The nomograms allow optimization and pricing to two significant figures. If in a particular case, greater accuracy is desired, this can be achieved by solving the above three equations directly. The development of these equations along with graphs for their solutions are given in the Appendix.

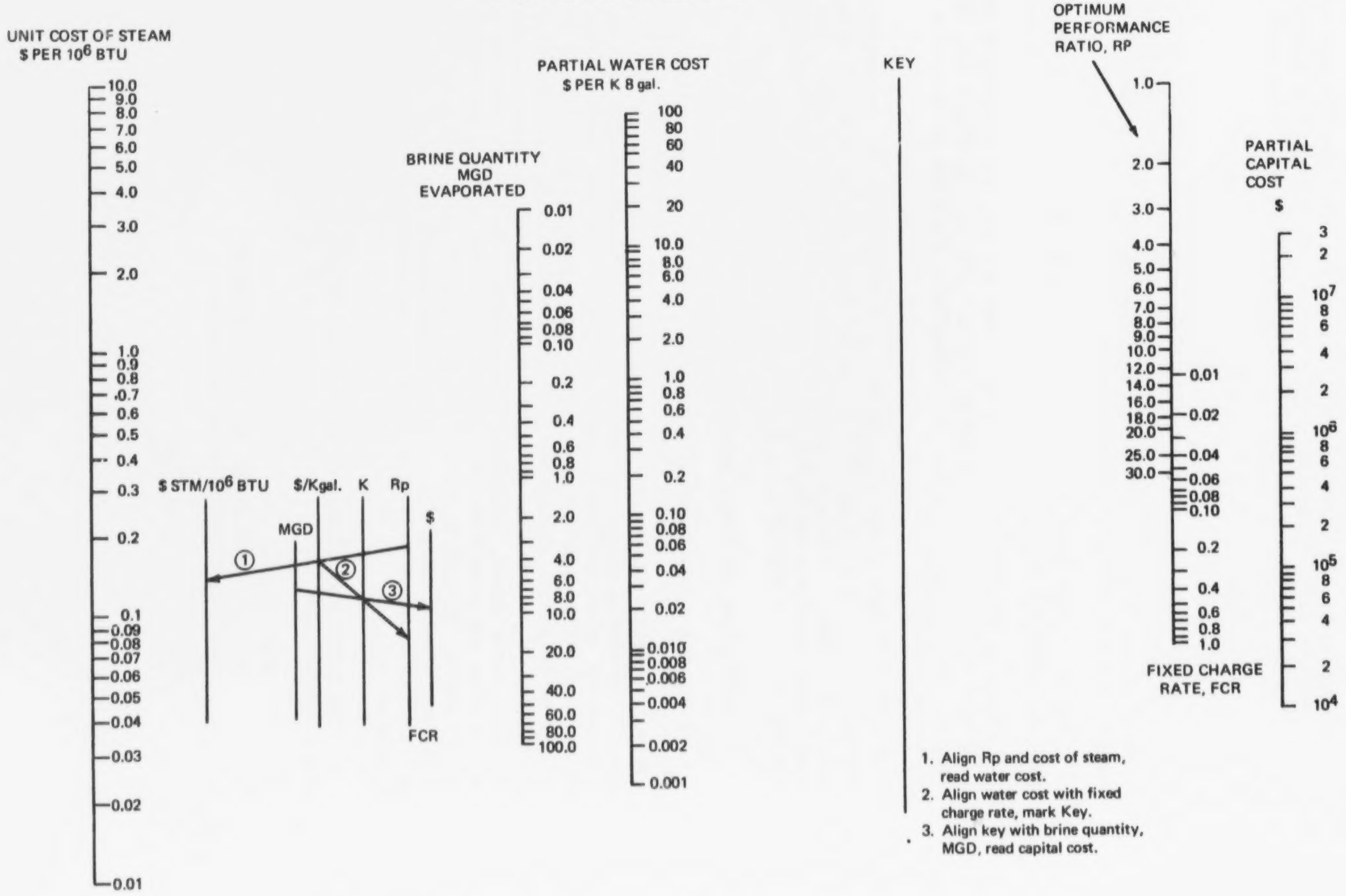
Determining the Total Cost of Water Evaporated

Since the water evaporated is also condensed by the multistage flash process, the cost of water evaporated is also the cost of product water. In municipal wastewater treatment, the brine fed to the evaporator will probably contain proteins that will decompose upon heating into ammonia and other odor-producing products. Therefore, this product water will be of little value, except perhaps for irrigation -- unless it is subjected to some post-treatment -- a subject that is outside the scope of this study. If the water fed to the evaporator, however, is well water or river water, the product is potable. In either case the economics that are presented herein are valid. The optimum performance ratio charts, therefore, have a utility that is not limited to a municipal wastewater renovation scheme. These charts may be used for any non-scaling evaporator feed that can be concentrated up to 10% total dissolved solids.

The total cost of the water produced is the sum of all the partial costs $/Kgal. due to:

Base capital cost
Partial water costs from optimization
Pumping costs for water and brine
Fixed costs
Operating labor costs

PARTIAL WATER COST AND CAPITAL COSTS VS. RP AND STM COST, NOMOGRAM 2



1. Align Rp and cost of steam, read water cost.
2. Align water cost with fixed charge rate, mark Key.
3. Align key with brine quantity, MGD, read capital cost.

These different costs are handled as shown in the following example for a 2.5 mgd evaporator plant.

From Nomogram No. 1, at FCR = 0.10, and Steam Cost = \$0.863/mbtu

$$Rp = 18, \quad N = 71$$

From Nomogram No. 2, at Rp = 18 and Steam Cost = \$0.863,

Partial Water Cost = \$0.79/Kgal. and \$Cap. Cost = \$2.4 million

Total Water Cost (excluding operating labor) = Base Cost + Partial Cost +

Pumping Cost + Fixed Cost = .0894 + 0.79 + .017 + .02 = \$0.92/Kgal.

Conceptual Design of Multistage Flash Plant for Brine Disposal

Once the capacity of the evaporator plant has been established, and the optimum performance ratio has been determined by using the above procedures, the conceptual design can be started. The performance ratio and number of stages shown by the nomogram will be optimum provided that the flashdown of the plant is a total of 250° F, the mean overall heat transfer coefficient is 630, and the mean effective boiling point elevation, including all allowances, is 4.3° F. For other conditions and for crossflow condensers a reoptimization will have to be conducted.

Using the three evaporator plant sizes required by this study for concentrating the waste brine, (0.1 mgd, 1.0 mgd, and 10.0 mgd) and for steam costing 46¢/mBtu (corresponding to natural gas at 35¢/mBtu), together with an annual fixed charge rate of 10%, optimum performance ratios and optimum numbers of stages have been determined, using the nomograms. These values have been fed into a computer using the Oak Ridge ORSEF code.* This program completely designs the water plant and prints out all of the design information, so that drawings can be made and detailed. Rectangular, multi-level modules are assumed for the cost calculations in the computer program, but the engineering information generated can equally well be applied to cylindrical vessels, and was done in a prior study (Bechtel '66).

Block diagrams (Figures 8 through 11) showing the water plant connections to the boiler and associated steam turbine drives for the water pumps are presented on the following pages. These differ from a seawater conversion plant in one important respect: the cooling of the multistage flash evaporator plant is done, not with seawater, but with renovated municipal wastewater, either preceding or following the quaternary electrodialysis step. This procedure, if used on the feed stream of the electrodialysis unit prior to acidification would raise the feed to 100° F, and the higher temperature enhances the operation of the electrodialyzer. The feed water

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The ORSEF code approximates the cost of operating labor as being:
Operating Labor, \$/Kgal. = $0.351\ (MGD)^{-.675}$

*See Appendix for computer printouts.

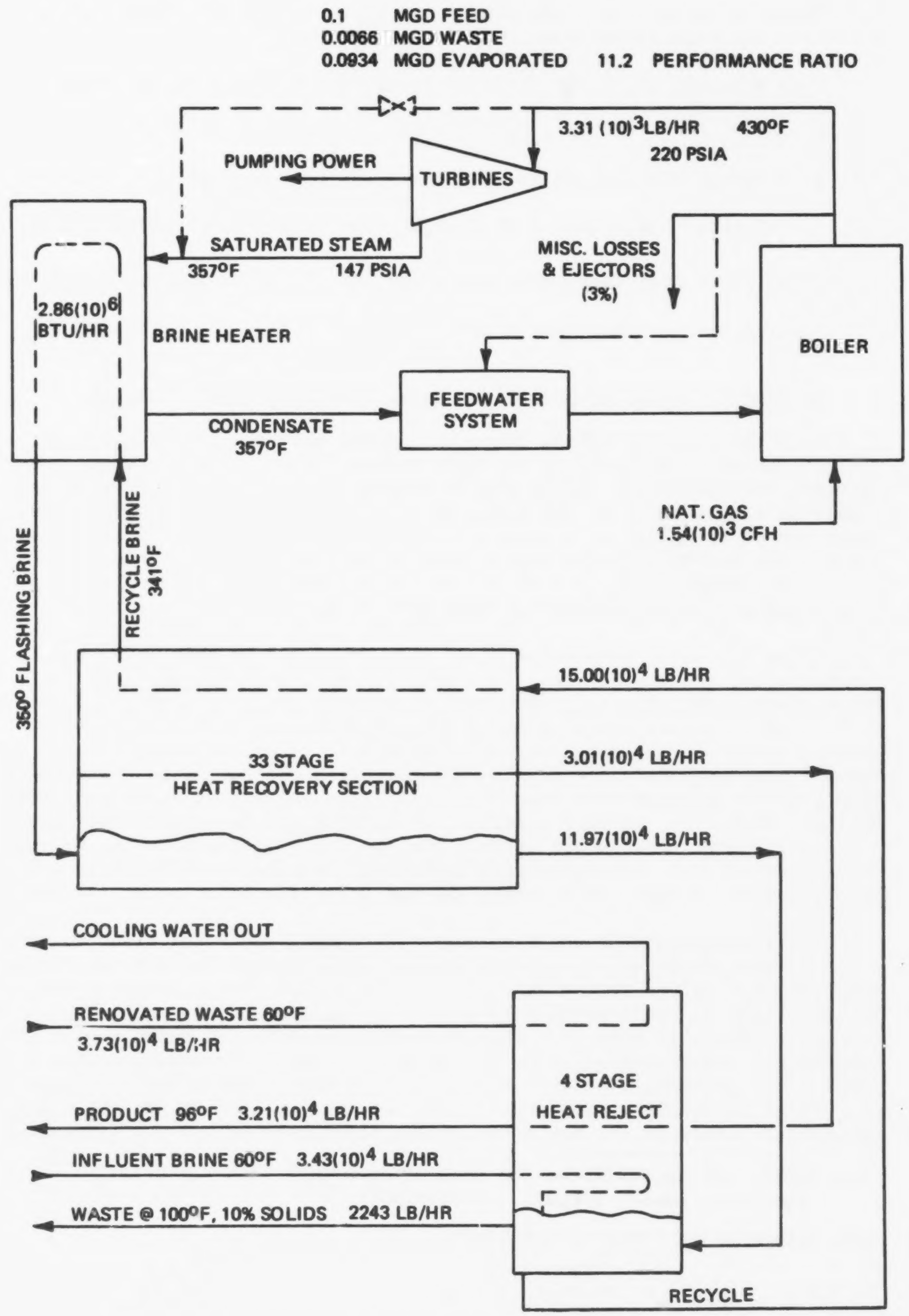


FIGURE 8. MULTISTAGE FLASH PRECONCENTRATOR – 0.1 MGD FEED

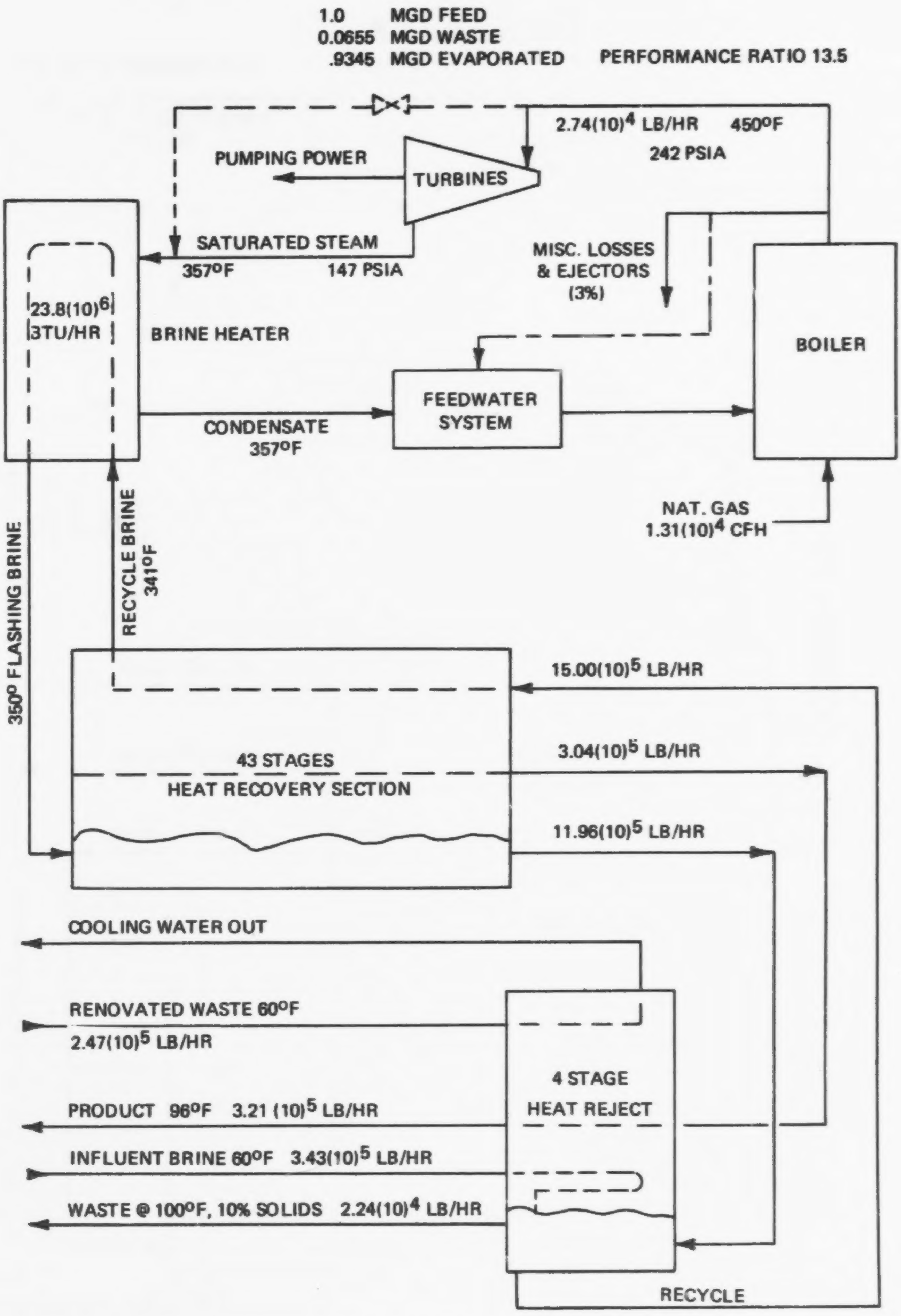


FIGURE 9. MULTISTAGE FLASH PRECONCENTRATOR - 1.0 MGD FEED

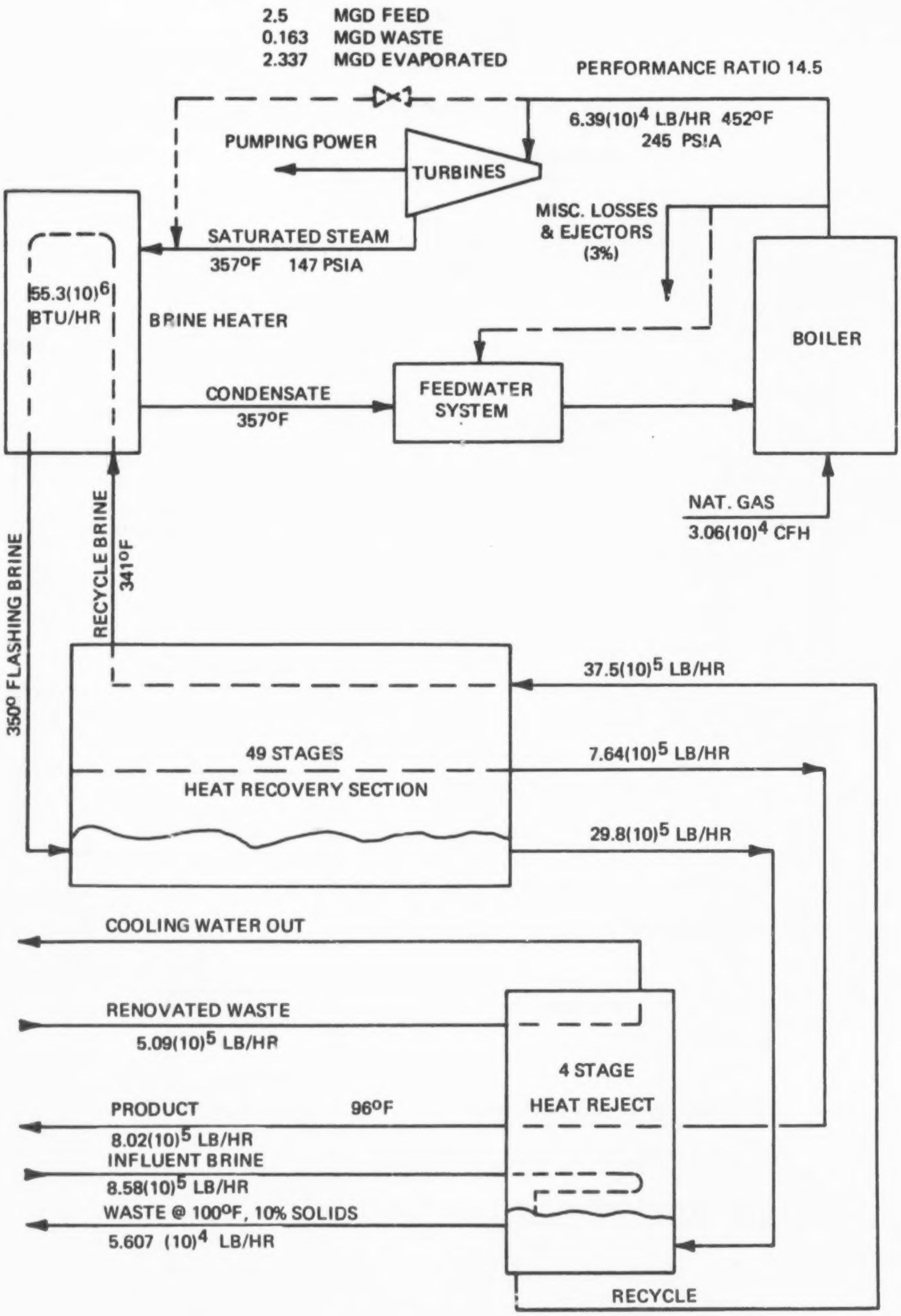


FIGURE 10. MULTISTAGE FLASH PRECONCENTRATOR - 2.5 MGD FEED

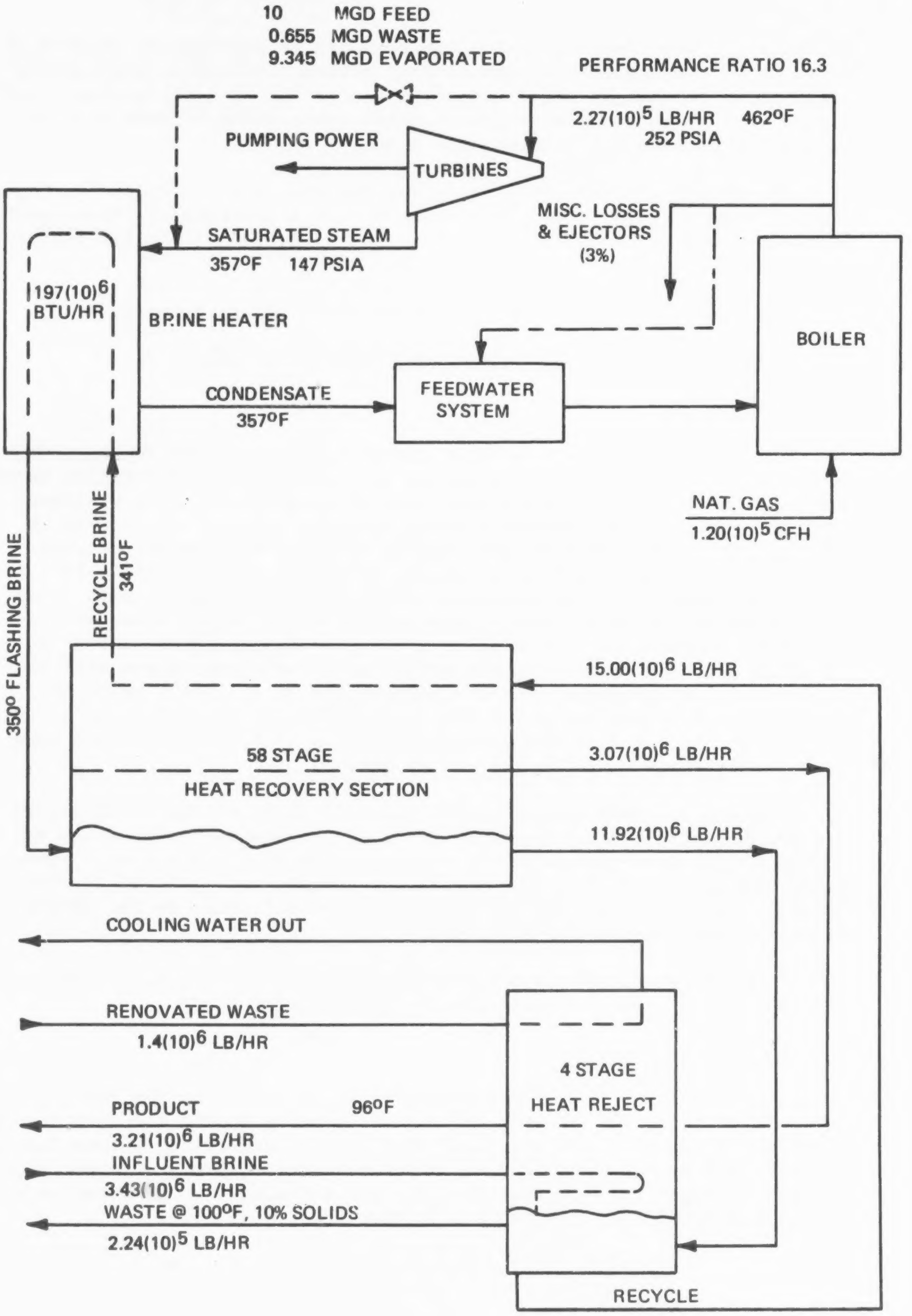


FIGURE 11. MULTISTAGE FLASH PRECONCENTRATOR - 10 MGD FEED

in the evaporator heat-rejection tubes, would be adjusted to about 8 pH so as to be non-aggressive, and so that savings in bundle construction costs could be made by using non-exotic materials of construction. Additional savings in evaporator construction costs could be made by eliminating the demisters throughout the plant.

The details of the plant equipment and the layout, except for the smaller size vessels, would be similar to those developed by Fluor and Bechtel, (1966). The OSW universal plant is also quite similar (Burns and Roe, '68).

The design data for MSF Evaporators are summarized in Table 4.

MULTISTAGE FLASH EVAPORATION WITH ION EXCHANGE PRETREATMENT

Introduction

The previous sections of this report have considered the economics of using multistage flash evaporation as a preferred concentrating means for brine from the electrodialysis step of an advanced waste treatment process. In such an evaporator plant, maximum economy is achieved by running the brine heater at the maximum temperature possible without scaling. By using cold lime treatment preceding the electrodialysis step, the hardness of the secondary sewage effluent can, in many cases, be reduced to the point where concentration to 10% total dissolved solids is possible without scaling a brine heater operated at 350° F. In other cases, a post-treatment of the electrodialysis brine will be necessary. In any event, the concentrate stream will have to be adjusted for pH before being fed into an evaporator. A typical pH of the concentrate stream from electrodialysis units is 4.0. This would have to be neutralized to a pH of 7.0 with soda ash.

One way to avoid costly post-treatment of the effluent brine from a conventional electrodialyzer would be to use sodium-cycle cation exchange on the feed stream to the evaporator, with the evaporator waste brine stream itself used as a regenerant for the ion-exchange columns. Thereby, the principal cost of the post-treatment would be the investment cost for the ion-exchange unit.

The operating costs for ion-exchange post-treatment at 0.1 mgd, 1.0 mgd, and 10 mgd are developed in the following pages.

Objective

The objective of this investigation is to determine the feasibility of ion exchanging the calcium and magnesium salts in the electrodialysis brine concentrate to sodium salts. The brine concentrate is then further concentrated by evaporation. The brine concentrate from the evaporator is used to regenerate the sodium cycle ion exchanger. Concentration to saturation with sodium chloride is theoretically possible.

TABLE 4

SUMMARY OF MSFE COMPUTER OUTPUT

Based on following Design Criteria:

Feed	7000 mg/1
Concentration Ratio	14.3
Heat Rejection Stages	4
Recovery Ratio	0.214 lb Product/lb In-Tube Brine
Steam Saturated At	357°
Brine, Maximum Temperature	350°
Blowdown Temperature	100°
Product Temperature	97.4°
Tubing Outside Diameter	0.75 in.
Tubing Wall Thickness	0.035 in.
Tubing Material	Cu-Ni
Fouling Factor	0.0004
Average Overall U	630 Btu/ft^2 hr °F
In-Tube Brine Velocity	5.25 ft/sec
Cooling Water Temperature	60°

Feed Capacity; mgd	0.1	1.0	10.0
Product Water; mgd	0.09	0.93	9.30
Optimum Performance Ratio	11.2	13.5	16.3
Steam Flow: lbs/hr	3,300	27,400	227,000
Pump Power Required; MW	0.03	0.31	3.23
Number of Evaporator Stages	33	43	58
Number of Trains	1	1	4
Brine Heater Surface Area; ft^2	266	2,380	21,100
Evaporator Surface Area; ft^2	3,750	47,000	596,000
Heat Rejection Surface Area; ft^2	252	2,250	19,900
Number of Tubes; Brine Heater	53	532	5,319
Number of Tubes; Evaporator	53	527	5,276
Number of Tubes; Heat Rejection	24	199	1,630

Discussion

Technical Approach: The electrodialysis process uses an electric current to separate cationic and anionic components from the waste water being renovated. Membranes allow the ions to pass from a dilute solution on one side of the membrane to a concentrated solution on the other side with a pH of the concentrated solution of about 4.0.

The total dissolved solids of the electrodialysis concentrate solution consists of 90% sodium salts and 10% calcium-magnesium salts. If the calcium-magnesium salts are exchanged to sodium salts, the resultant water stream is favorable to an evaporation process and the dissolved solids are further concentrated. The waste stream from the evaporator, using multistage flash, is a 10% brine solution and can be used as the regenerant for the sodium cycle ion exchanger.

The composition of the electrodialysis brine concentrate is shown in Table 5, Column A. The concentrate contains 700 mg/l Ca and Mg as $CaCO_3$ and 6,300 mg/l Na as $CaCO_3$. The anions are 2,550 mg/l Cl as $CaCO_3$ and 4,450 mg/l SO_4 as $CaCO_3$. In the sodium cycle ion exchange process the resin is in the sodium form. The calcium and magnesium ions in the water are exchanged to sodium ions, producing a water containing sodium chloride and sodium sulfate.

The effluent from the ion exchanger feeds to an evaporator. The evaporator produces a distillate yield of 93% of the feed stream. The waste stream containing the sodium salts is 7% of the feed stream.

The evaporator waste stream is composed of 10% of sodium chloride and sulfate salts. A portion of this stream could be used for regenerating the sodium cycle ion exchange unit. The remainder of the evaporator waste stream would be used to rinse the resins before being discharged to waste. The spent regenerant solution containing calcium and magnesium salts is also discharged to waste.

Results: The analysis of the electrodialysis brine concentrate fed to the ion exchange column shown in Column A. The concentrate contains 700 mg/l calcium and magnesium (as $CaCO_3$) and 6300 mg/l sodium (as $CaCO_3$). At a water flux of 5 gpm per square foot, the average Ca-Mg leakage in the effluent leaving the ion exchanger will be 120 mg/l Ca-MgH (as $CaCO_3$). This was determined by extrapolating a hardness versus total dissolved solids curve from 5,000 mg/l tds to 7,000 mg/l tds. The expected resin exchange capacity is 22 kilograins per cubic foot (as $CaCO_3$) when regenerated with 15 pounds NaCl per cubic foot resin. This value was obtained from an extrapolated curve of exchange capacity versus total dissolved solids.

For a one million gallon per day treatment plant 27,900 pounds sodium chloride per day is required. The waste stream from the evaporator consists of 10% salts or 58,400 pounds per day, sodium chloride-sulfate. Approximately 48% of this salt will be consumed in the regeneration of the ion exchanger. The remaining solution will be used for rinsing and then discharged to waste. The spent regeneration solution containing calcium and magnesium salts will be discharged to waste also. To avoid dilution of the waste, no dilute brine will be used for rinsing.

TABLE 5

WATER ANALYSIS-ULTIMATE DISPOSAL-ELECTRODIALYSIS BRINE

Constituents	"A"	"B"	"C"	"D"
mg/l as $CaCO_3$	E.D. Brine Conc.	Na Cycle I.X.	Evapor. Distillate	Evapor. Blowdown
Cations				
Calcium	700	120	0	1,716
Magnesium				
Sodium	6,300	6,880	0	98,384
Ammonia, Free	1,000	1,000	1,000	-
Total Cations	7,000	7,000	-	100,100
Anions				
Bicarbonate	0	0	0	0
Carbonate	0	0	0	0
Hydroxide	0	0	0	0
Chloride	2,550	2,550	0	36,465
Sulfate	4,450	4,450	0	63,635
Phosphate	0	0	0	0
Total Anions	7,000	7,000	-	100,100
Other Analysis				
Total Hardness	700	120		1,716
Total Alkalinity	0	0		0
P Alkalinity	0	0		0
Free Carbon Dioxide	0	0		0
Total Dissolved Solids	7,000	7,000		100,000

Conclusions and Summary

A one hundred thousand gallon per day sodium cycle treatment plant would cost $15,000, and the installation would cost approximately $7,000-$8,000. The softening cost with 7% FCR would be 32.7¢ per 1,000 gallons.

The capital cost of a sodium cycle ion exchange plant to produce one million gallons softened water per day is $50,000 and $25,000 to install the plant. The softening cost with 7% FCR is 5.0¢ per 1,000 gallons water.

The capital cost of a 10 million gallon per day plant is $300,000, and the installation cost is $150,000. The softening cost with 7% FCR is 1.75¢ per 1,000 gallons water softened.

These capital and operating costs are high for small size plants but become reasonable for larger sizes. Cost breakdowns are as follows:

Plant Size, MGD.	0.1	1.0	10.0
Capital Cost, $	15,000	50,000	300,000
Installation Cost, $	7,500	25,000	150,000
Total Cost, $	22,500	75,000	450,000
At 5% FCR $/Kgal.	.0343	.0114	.00683
At 10% FCR $/Kgal.	.0685	.0228	.0137
Power Cost @ 10 mills, plus Operating Labor $/Kgal.	.256	0.0264	0.0035
Maintenance and Repair including Resin Loss @ 5% Cap. $/Kgal.	0.0228	0.0076	0.0045
Total Capital and Operating Cost per K/gal. Softened			
At 5% FCR $/Kgal.	0.3131	0.0454	0.0148
At 10% FCR $/Kgal.	0.3473	0.0568	0.0217

VAPOR COMPRESSION AND MULTIEFFECT EVAPORATION

Introduction

For concentrating brines beyond 10% tds, vapor compression or multieffect evaporators are generally used, and different methods are employed to handle salt and scale depositions. Practical criteria, rather than optimization criteria govern. Unit costs are functions of concentration and capacity. Concentration to saturation, instead of dryness is generally more feasible. Capital costs and operating costs per 1000 gallons (Kgal.) of water evaporated have been developed. These costs are directly additive with conveyance costs to the nearest ultimate disposal site for the residue to give the total cost for disposal in suitable units.

Objective

To concentrate beyond 10% solids, multieffect evaporators, when properly designed, have an economic advantage both over single effect evaporators and over multistage flash evaporators. The latter, because they are not primarily designed as concentrators, are seldom used above 10% solids. Of the various multieffect arrangements, vapor compression evaporators, which multieffect by forced recirculation, are particularly suitable for concentrating brine to saturation for disposal. The purpose here is to concentrate the brine to a minimum volume for disposal without going through an expensive crystallization step, involving special heat exchangers. Trouble in the evaporator can be avoided if concentration does not proceed beyond saturation, which for sodium chloride is about 25% by weight.

Procedure

The flow diagram (Figure 12) shows how vapor compression operates. The feed stream first enters a liquid-liquid heat exchanger, and is preheated to the operating temperature of 228° by cooling the product water and the saturated brine blowdown. The feed stream is assumed to be the deaerated brine blowdown from a preceding multistage flash evaporator and to contain 10% total dissolved solids. It is concentrated by vapor compression to 25% solids.

The feed stream next mixes with recycled brine from the evaporator and enters the recirculating pump. It is pumped upward through a vertical condenser and boils against the tube walls while condensing steam from the vapor compressor, evolving thereby an equal quantity of steam to that condensed.

The steam-brine mixture next enters the evaporator dome, where separation of the two phases occurs. Due to the boiling point elevation of saturated brine, the steam evolved is superheated about 16° F, at 14.7 psia and reaches the inlet of the vapor compressor in this condition. It is compressed isentropically to 342° F and then desuperheated to 238° F at 24 psia so as to provide a 10° thermal driving force across

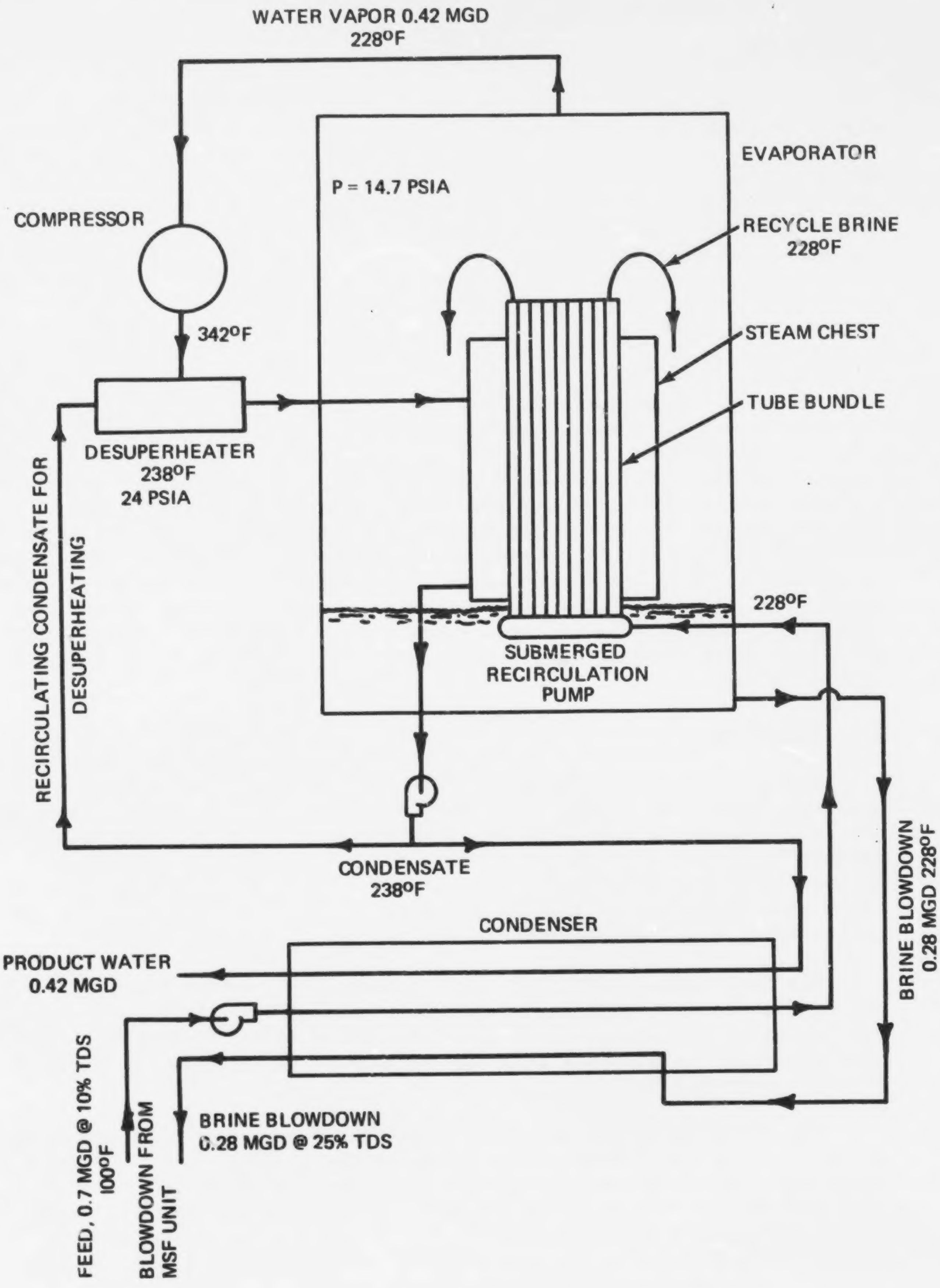


FIGURE 12. VAPOR COMPRESSION PLANT FLOW DIAGRAM

the condenser. The condenser must be provided with a deaerating section (not shown) to remove any non-condensables not already removed by the preceding multistage flash evaporator. The residual brine collects in the bottom of the evaporator, from which a portion is blown down at 25% solids, while the remainder mixes with the preheated feed stream to complete the cycle.

Discussion

The fact that the feed stream consists of the blowdown stream from a preceding multistage flash evaporator means that the feed stream is almost completely deaerated before entering the vapor compression unit. Since non-condensables constitute one of the principal irreversibilities of the vapor compression cycle, a completely deaerated feed can be more economically processed by vapor compression than by multiple effect, at ordinary power costs. Moreover, the use of electrical energy saves the capital cost of the boiler plant otherwise required with multieffect evaporators. With practical temperature approaches in the heat exchangers, the total energy consumption of the compressor, at 70% overall efficiency will run about 132 Kwhr/Kgal. Pumping energy will run about 8 Kwhr/Kgal., so that the total energy consumption of the evaporator unit will be 140 Kwhr/Kgal.

A major disadvantage of vapor compression is the large amount of operating and maintenance labor necessary to keep the compressor blades in balance and the tube surfaces free from deposits, when operating at high solids concentrations. For the intermediate size plant, we have allowed three man-days per month to cover part-time operating labor and periodic cleaning of the tubes. For a small size unit such as this, when considered in relation to the size of the complete waste renovation plant of which it is a part, this proration of man-hours is probably sufficient.

Conclusions and Summary

As an example of the total operating costs for concentrating 10% brine to 25% brine for 1.0, 10.0 and 100 mgd waste water renovation facilities, with an annual fixed charge rate of 10% and a power rate of 12 mills per kilowatt-hour, the estimated costs from the accompanying graph (Figure 13) are respectively:

Water Evaporated, Kgal./day	4.2	42	420
Cost in $/Kgal. Evaporated	$5.48	$3.31	$2.44

The components of cost making up the total cost given in Figure 13 are presented in Table 6. The information given in Table 6 was taken from: Dodge & Eshaya; Chambers & Larsen, '60; and OSW '68.

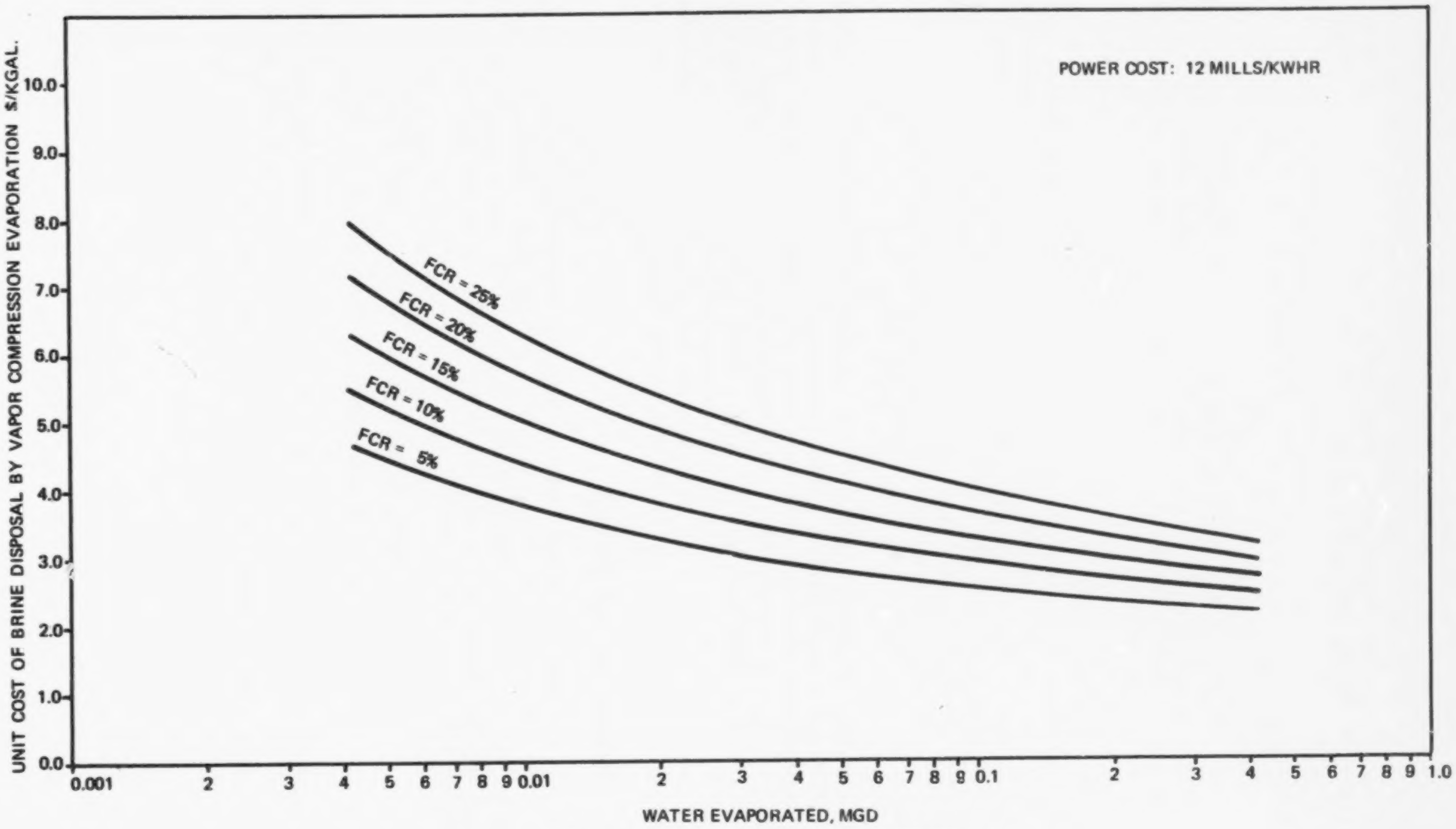


FIGURE 13. VAPOR COMPRESSION EVAPORATION

TABLE 6

VAPOR COMPRESSION EVAPORATOR COSTS

Feed mgd @ 10% Solids	0.7	0.07	0.007	
Product H_2O Gal./Day	420,000	42,000	4,200	
Brine to Disposal @ 25% Solids	280,000	28,000	2,800	
Capital Cost $	714,000	126,000	22,650	
$/Daily Gallon	1.70	3.00	5.40	
$/Annual Kgal.	5.17	9.13	16.40	
$/Kgal. @ FCR = .05	0.259	0.457	0.82	A
FCR = .10	0.517	0.913	1.64	
FCR = .15	0.775	1.37	2.46	
FCR = .20	1.034	1.826	3.28	
FCR = .25	1.293	2.28	4.10	
Power Cost @ $\frac{140\ \text{Kwhr}}{\text{Kgal.}}$				
$/Kgal. @ 5 mills		.70		B
@ 10 mills		1.40		
@ 12 mills		1.68		
@ 15 mills		2.10		
@ 20 mills		2.80		
@ 25 mills		3.50		
O. & M. Labor $/Kgal.	0.24	0.72	2.16	C

Total Cost $/Kgal. Evap. = A + B + C

SUBMERGED COMBUSTION

Introduction

The brine effluent of the multistage flash evaporator (MSFE) discussed previously in this report, is about 100 times as concentrated as the original secondary sewage treatment effluent. However, even after this degree of concentration has occurred, the volume of the brine is still too large to be economically transported to an ultimate disposal site. Where solar evaporation ponds cannot be used, multistage flash evaporation is the most economical alternative for concentrating brine to 7%. For concentrating the brine further, another method which may be more economical is evaporation by submerged combustion. This technique can be used to increase brine concentration to 50 or 60% tds, a concentration which is more than sufficient for economical transportation to an ultimate disposal site.

Objective

The objective of this section is to examine the use of submerged combustion evaporation to concentrate multistage flash evaporation blowdown from seven to fifty percent tds. Approximate costs have been determined for 10,000 gpd and 1,000,000 gpd SCE plants.

Discussion

Submerged combustion evaporation (SCE) has two principal economic advantages for the concentration of MSFE brine effluent. The first is low capital cost which, for a 1.0 mgd plant large enough for a sewage effluent of 100 mgd, is $1,520,000 (see Table 7). The second advantage is low maintenance costs. The addition of an SCE plant to an MSFE plant, ten times as large, would probably not require any additional personnel. The low capital and maintenance costs are largely a result of the absence of condensers and heat exchangers. At an annual fixed charge rate of 10%, total costs for the 1.0 mgd plant are $4.90/Kgal. evaporated and $0.0435 per/Kgal. of sewage plant effluent (see Table 8). Costs for the 0.1 mgd plant are just slightly higher and total costs for the 10,000 gpd plant, large enough for 1.0 mgd of sewage plant effluent, are $6.58/Kgal. evaporated and $0.0586/Kgal. sewage plant effluent. In this process no heat recovery is attempted, which leads to high operating costs (see Table 7), accounting for from 50 to 70% of the total submerged combustion evaporation cost for 10,000 gpd plant, and 75 to 90% of the costs for the two larger plants (see Table 8). Operating costs could be reduced by using the vented gases for preheating the feed brine, but the adverse effect on capital costs caused by introducing heat exchangers would probably outweigh the operating cost reduction and result in an increased total cost of the process. For economical transportation to an ultimate disposal site, the effluent of an MSFE must be reduced to about one-seventh its initial volume. This means about 92% of the water in the MSFE brine must be evaporated by submerged combustion.

TABLE 7

DATA ON CONCENTRATING INORGANIC WASTE STREAMS FOR 7 TO 50 PERCENT SOLIDS BY SCE

	CASE I	CASE II	CASE III
MGD of Sec. Sewage Effluent	1	10	100
Percent Solids in Feed to Sub. Comb. Ev.	7%	7%	7%
Feed Rate of 7% Brine Stream (GPD)	10,000	100,000	1,000,000
Feed Rate of 7% Brine Stream (GPH)	417	4,170	41,700
Feed Rate of 7% Brine Stream (lb/hr)	3,480	34,800	348,000
Solids Rate (Dry Basis) (lb/hr)	243	2,430	24,300
Product Rate at 50% Solids (lb/hr)	486	4,860	48,600
Water Evaporated (lb/hr)	2,994	29,940	299,400
Firing Rate (10^6 Btu/hr)	4	40	400
Number of Units	1	1	8
Fuel Cost at $\$0.35/10^6$ Btu (\$/hr)	\$ 1.40	\$ 14.00	\$ 140.00
Electrical Cost at 12 mills/kwh (\$/hr)	\$ 0.27	\$ 1.80	\$ 18.00
Total Operating Cost Less Labor (\$/hr)	\$ 1.67	\$ 15.80	\$ 158.00
Total Operating Cost Less Labor (\$/1000 gal. evaporated)	\$ 4.64	\$ 4.37	\$ 4.37
Approximate Capital Cost Prefabricated with Scrubber and Installed	\$55,000	\$163,000	\$1,520,000
Annual Capital Cost at 5%/yr Fixed Charge Rate (\$/yr)	\$ 2,750	\$ 8,150	\$ 76,000
Annual Capital Cost at 10%/yr Fixed Charge Rate (\$/yr)	\$ 5,500	\$ 16,300	\$ 152,000
Annual Capital Cost at 12.5%/yr Fixed Charge Rate (\$/yr)	\$ 6,875	\$ 20,375	\$ 190,000
Annual Capital Cost at 15%/yr Fixed Charge Rate (\$/yr)	\$ 8,250	\$ 24,450	\$ 228,000
Annual Capital Cost at 20%/yr Fixed Charge Rate (\$/yr)	\$11,000	\$ 32,600	\$ 304,000
Annual Capital Cost at 25%/yr Fixed Charge Rate (\$/yr)	\$13,750	\$ 40,750	\$ 380,000

TABLE 8

UNIT COSTS FOR CONCENTRATING INORGANIC WASTE STREAMS FROM 7 TO 50 PERCENT BY SCE (90% LOAD FACTOR)

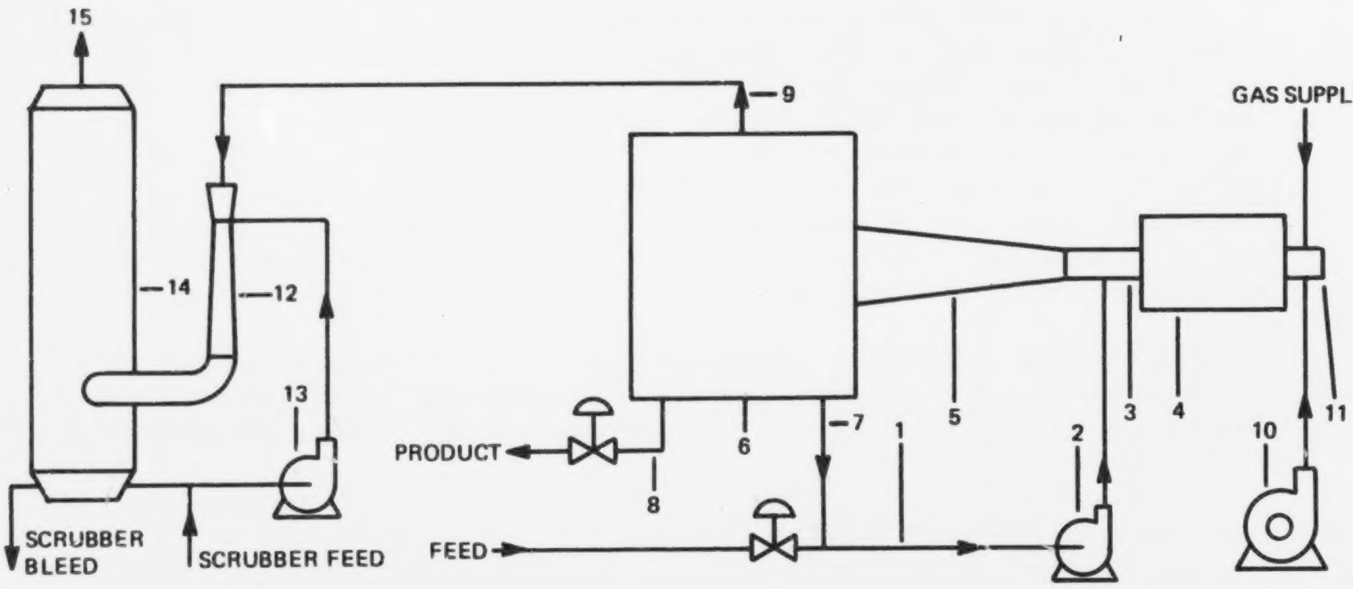
	CASE I	CASE II	CASE III
Sec. Sewage Effluent (MGD)	1	10	100
Feed Rate 7% Brine Stream (GPD)	10,000	100,000	1,000,000
Capital Cost at 5%/yr FCR ($/1000 gal. evaporated)	$ 0.98	$ 0.29	$ 0.27
Capital Cost at 10%/yr FCR ($/1000 gal. evaporated)	1.96	0.58	0.53
Capital Cost at 12.5%/yr FCR ($/1000 gal. evaporated)	2.45	0.71	0.63
Capital Cost at 15%/yr FCR ($/1000 gal. evaporated)	2.93	0.87	0.81
Capital Cost at 20%/yr FCR ($/1000 gal. evaporated)	3.92	1.16	1.07
Capital Cost at 25%/yr FCR ($/1000 gal. evaporated)	4.90	1.43	1.34
Total Cost at 5%/yr FCR ($/1000 gal. evaporated)	5.62	4.66	4.64
Total Cost at 10%/yr FCR ($/1000 gal. evaporated)	6.58	4.95	4.90
Total Cost at 12.5%/yr FCR ($/1000 gal. evaporated)	7.08	5.08	5.04
Total Cost at 15%/yr FCR ($/1000 gal. evaporated)	7.56	5.24	5.18
Total Cost at 20%/yr FCR ($/1000 gal. evaporated)	8.56	5.53	5.43
Total Cost at 25%/yr FCR ($/1000 gal. evaporated)	9.52	5.80	5.71
Total Cost of SCE 5%/yr FCR ($/1000 gal. of Sec. Effluent)	0.0500	0.0414	0.0412
Total Cost of SCE 10%/yr FCR ($/1000 gal. of Sec. Effluent)	0.0586	0.0439	0.0435
Total Cost of SCE 12.5%/yr FCR ($/1000 gal. of Sec. Effluent)	0.0629	0.0452	0.0448
Total Cost of SCE 15%/yr FCR ($/1000 gal. of Sec. Effluent)	0.0672	0.0466	0.0461
Total Cost of SCE 20%/yr FCR ($/1000 gal. of Sec. Effluent)	0.0760	0.0492	0.0483
Total Cost of SCE 25%/yr FCR ($/1000 gal. of Sec. Effluent)	0.0846	0.0515	0.0508

The submerged combustion evaporator operates as a direct contact evaporator (see Figure 14). Fuel gas and air are mixed and then fired by a high energy release burner into a combustion chamber. Hot combustion products from this chamber are exhausted through a small port. The liquid to be evaporated, in this case MSFE effluent mixed with recycled product (the product is already at 50% tds), is injected at a controlled rate from an annular feed chamber jacket in the exhaust port of the combustion chamber. A complete exchange of heat takes place instantly as the liquid stream enters the exhaust gas stream.

The heated product is separated from the quenched gas stream in a cyclonic separator. Spent gases are exhausted and the product is collected and drawn off for ultimate disposal and for recycling with fresh MSFE effluent.

Conclusions and Summary

Estimates of costs have been prepared for submerged combustion evaporators for three capacities, 10,000 gpd, 0.1 mgd, and 1.0 mgd (see Tables 7 and 8). The estimates are predicated on a natural gas price of $.35 per million Btu, power costs of 12 mills per kwh, and no maintenance cost. Capital costs for units to be used with 10,000 gpd, 0.1 mgd, and 1.0 mgd plants, including prefabrication with a scrubber and installation, have been estimated at $55,000, $163,000, and $1,520,000, respectively. Operating costs per thousand gallons evaporated have been estimated at $4.64, $4.37, and $4.37, respectively. The total unit cost for submerged combustion evaporation will, of course, depend on the fixed charge rate. The total costs for submerged combustion evaporation per thousand gallons evaporated and per thousand gallons sewage plant effluent are given for several fixed charge rates in Table 8 and Figure 15 respectively. For a 1.0 mgd plant with a 10% FCR, the cost is $4.90/Kgal. evaporated. Because of the prior combined concentrating effects of the electrodialysis plant and MSFE, there is only one gallon evaporated by SCE for each 116.28 gallons of sewage plant effluent. The resulting cost for concentration by SCE is $0.0435/Kgal. of sewage plant effluent. Although the above costing procedures have been based upon evaporating a 7% brine feed, the costs per 1,000 gallons of water evaporated should remain essentially the same if a 10% brine feed concentration is used.



1 CIRCULATING SYSTEM
2 CIRCULATING PUMP
3 FEED CHAMBER
4 COMBUSTION CHAMBER
5 VENTURI
6 SEPARATOR
7 RECIRCULATION
8 PRODUCT DISCHARGE
9 EXHAUST DUCT
10 COMBUSTION AIR BLOWER
11 BURNER
12 MODIFIED VENTURI SCRUBBER
13 SCRUBBER CIRCULATING PUMP
14 SCRUBBER SEPARATOR SECTION
15 EXHAUST

DESCRIPTION OF OPERATION

Feed solution is introduced into the circulation system (1) and pumped (2) to the feed chamber (3) where it meets the products of combustion jetting from the combustion chamber (4). The combustion products and the solution mix in the Venturi (5) and enter the separator (6) where the gases and the solution are separated. Some of the solution is recirculated (7) by the circulation pump (2) as product is drawn off (8) and the gases are vented (9). The liquid level in the separator is maintained by a level control which regulates the product valve. The feed valve is signalled from the specific gravity measured ahead of the feed chamber.

Combustion air is supplied by a positive displacement blower (10) equipped with a filter silencer and a snubber. Fuel gas is regulated by the fuel gas controls and ignited in the combustion chamber in a burner (11).

The stack gases from the separator are scrubbed in the modified venturi scrubber (12), fed with weak liquor by the scrubber circulating pump (13). The scrubbed gases are separated from the liquor in the separator section (14) and discharged (15).

FIGURE 14. TYPICAL FLOW DIAGRAM
SUBMERGED COMBUSTION EVAPORATION

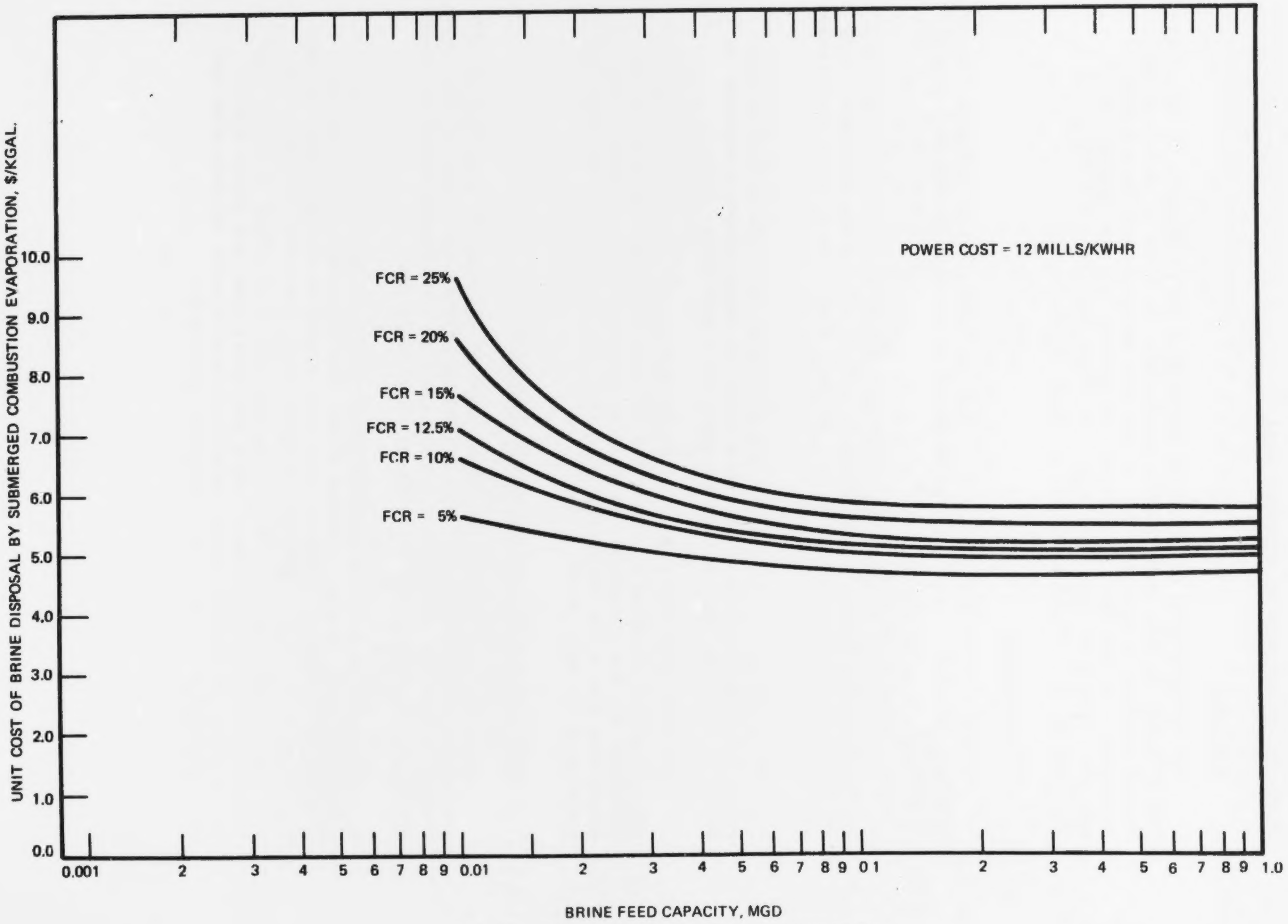


FIGURE 15. SUBMERGED COMBUSTION EVAPORATION

PIPELINE CONVEYANCE TO DISPOSAL AREA

Abstract

The optimization of pipeline conveyance costs depends on the pumping velocity as the optimization parameter. As this is a multidimensional problem, a nomogram and a computer program have been developed to express the results. Parametric equations for the capital costs and the conveyance costs per Kgal. of brine have been developed as a function of the distance to the ultimate disposal site.

Introduction

Conveyance by pipeline is an integral part of most ultimate waste disposal methods. It is also considered as a complete disposal method in itself.

The effluent brine of the treatment facility is transferred via a pipeline to the nearest site where it can be ultimately disposed (such as the sea, a salt lake, an injection well field, or a lined evaporation solar pond).

Our purpose is to evaluate the unit cost for this method of disposal and compare it with the unit costs of other disposal methods in order to find the best method to apply to a particular site.

The unit cost for such a disposal method is, in fact, the unit cost of water conveyance, since the effect of changes in density and viscosity of the wastes on conveyance cost is minor. (Koenig, '66, shows that for a 55% concentrated brine, the increase in the cost of conveyance is only about 3%.)

Cost Analysis of Conveyance by Pipeline

Costs of Pipelines: The total cost of disposal by pipeline is the sum of the partial costs of the following components: cost of the pipe (installation and material and OMR), cost of the pumping stations (pumps installed and OMR), and the cost of electrical power.

These partial costs are functions of the diameter of the conveyor. A designer may, in fact, choose almost any arbitrary pipe diameter. The use of a smaller diameter pipe will result in a lower initial capital investment, but a greater pressure drop, requiring larger pumps and higher power costs. Similarly, the use of a larger diameter pipe results in higher investment costs but lower pump and power costs. Accordingly, for every given capacity there is some optimum diameter that minimizes the cost of conveyance (disposal).

Cost Optimization: Finding this optimum diameter has been the purpose of many previous studies. Burwell ('67) developed an analytical expression to represent the system. As a result, he was able to optimize his system analytically to obtain optimum pipe diameter and cost of conveyance for his particular set of economic parameters.

There are, however, some simplifications in the Burwell report. An effort has been made to eliminate them in this analysis. For example, in calculating the friction pressure drops for the flow in pipes, Burwell used the Hazen-Williams formula because of the difficulty of applying the more accurate Darcy's formula in a parametric study. The use of a digital computer to perform the calculation allows application of Darcy's formula. Thus, more accurate results for the pressure drop and for the optimum diameter and costs have been obtained in this study.

Another major improvement on Burwell's approach is the computer selection of the nominal, rather than the theoretical, pipe diameter for the optimized system. In his cost equation, Burwell uses continuous, rather than discrete, values of the pipe diameter, in optimizing his cost equation mathematically. The resulting diameter is, in most cases, a non-standard size. However, a standard-size diameter would have to be used in practice. Its use would result in an off-optimum, but practical minimum, disposal cost. This limitation was neglected by Burwell. Comparing the total conveyance costs for the two neighboring standard-size diameters of the mathematical optimum, the practical minimum is obtained in this study by choosing the one with the lower operating cost.

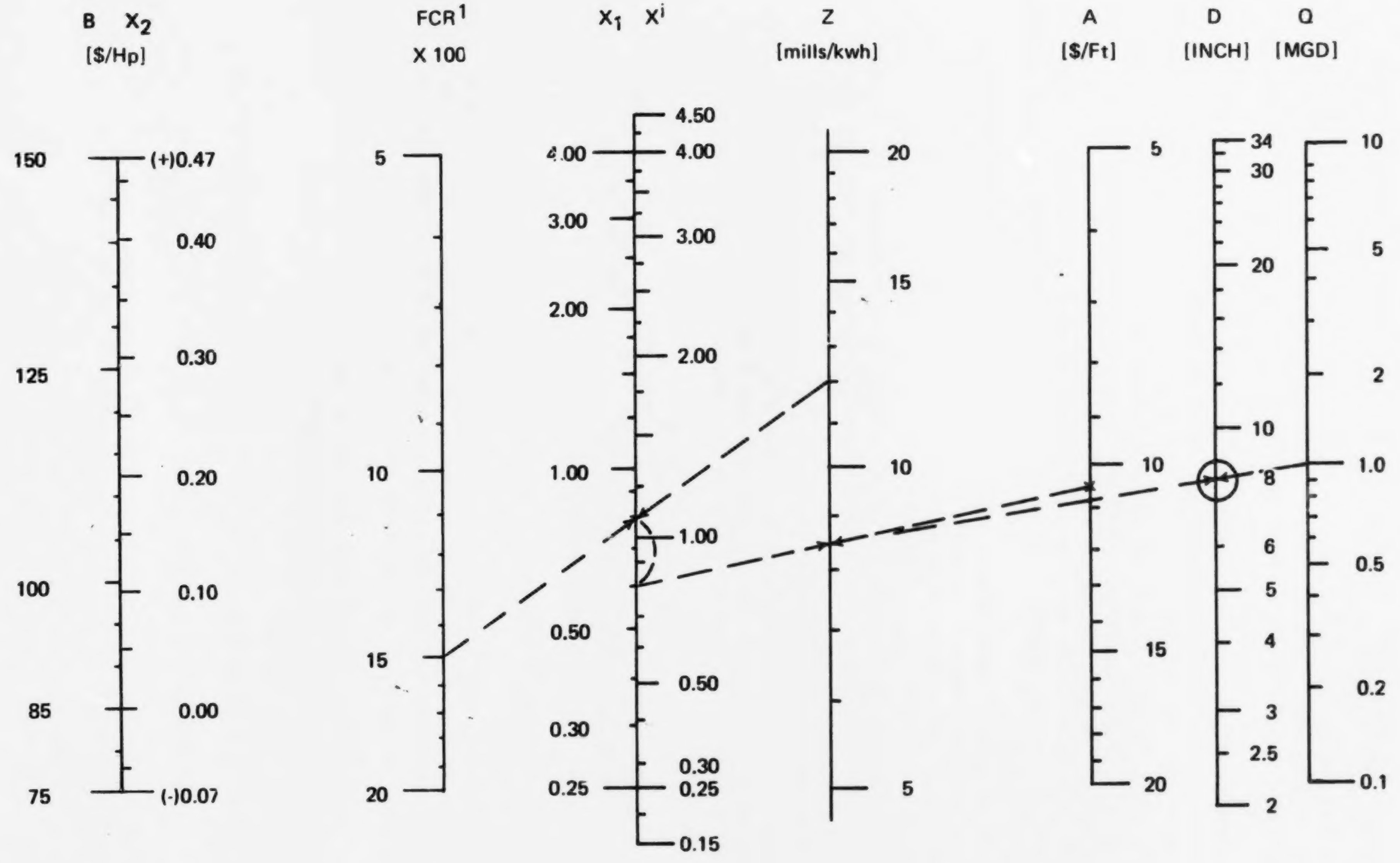
Nomogram: The cost of conveyance has been developed in terms of dollars paid for conveying 1,000 gallons a distance of 100 miles, or units of cents per kilo-gallon-miles.

The cost of conveyance is prorated to the capacity of the line on a yearly basis. It is the sum of the partial costs resulting from the cost of material and labor invested in the pipe, the cost of pumps required to overcome the pressure drop, the cost of maintenance of the pumping stations, and the cost of electrical energy required to drive the pumps. (Note: Only horizontal pipes have been considered.)

Upon this basis, the following equation has been developed for the optimum pipeline size. (See Appendix for mathematics.)

$$D^{6.29} = \frac{Q^3}{A}\left[1.508 \times 10^{-3} \times B + 0.719\ (Z/FCR')(10)^{-2}\right] = Q^3 \times A_3$$

The following nomogram (No. 3) solves this equation.



NOMOGRAM NO. 3 UNIT COST OF BRINE DISPOSAL BY PIPELINE CONVEYANCE, $/Kgal. X 100 MILES.

Use of Nomogram to Obtain Optimization Diameter

Required parameters:

FCR' = Fixed charge rate + 0.0025
Z = Power cost; (mills/kwh)
B = Pump cost; ($/HP)
A = Base cost = cost of a 12-in. diameter pipe,($) per linear foot

Using values of Z and FCR', find value of X_1, on respective scales.

Using value of B, find corresponding X_2.

$$X = X_1 + X_2$$

Using values of X' and A, find point on $K_{X'A}$ line (which coincides with Z line).

Using point on key line and desired capacity of pipeline as pivot points, find optimized' diameter (D).

Example:

$Z = 12$; $FCR' = (0.1475 + 0.0025) = 0.15$; $B = 85$; $A = 10.43$
Z and FRC' $X_1 = 0.80$
B $X_2 = 0$
$X' = 0.80$
X' and A point on key line
for Q = 1 mgd; D 8.0 in.
Use D = 8 inches

The numerical value of D is, in most cases, a non-standard diameter. In these cases the cost of conveyance must be evaluated for the two standard pipe sizes adjacent to the calculated value of D, and compared with them. The standard diameter with the lower cost is then selected.

Conclusions and Summary

Once the optimum pipe diameter has been selected the unit cost \$/Kgal. of conveyance per ℓ miles can be determined by use of the following cost equation:

$$CP_u = \left(\frac{.01\ell}{Q}\right) [1.447\ A(FCR')D^{1.29} + 2.738(10)^{-4}(FCR')R + (50f_d\ Q^3/D^5)$$

$$(5.628(10)^{-4}(FCR')B + 2.683(10)^{-3}Z)]$$

The pipeline sizes and costs appearing in this report were determined by the computer program appearing in the Appendix.

For the above case, with a friction factor of 0.02 and right-of-way cost of \$5000/mi.,the conveyance cost per 100 miles will run \$1.632/Kgal.

OTHER METHODS

Introduction

According to the Technical Approach, processes which are ancillary, such as pretreatments and post-treatments, and processes which appear to be of dubious feasibility are to be given merely cursory treatment.

Of these methods, Impoundment with Controlled Release of Brines is not applicable to the selected sites, and therefore has not been developed in detail; Brine Desulfation employs Barium, which is illegal for use in water supplies; Direct Contact Oil-Water Evaporation has serious technical difficulties, and has never been reduced to practice; Electrodialysis with ion-specific membranes turns out to be a most promising process, which however, is generally outside the scope of this study because it is part of the quaternary treatment process preceding the operations under study; therefore, in this study it has been considered as a post-treatment of AWT wastes, and not as a disposal step.

Impoundment and Controlled Release of Wastes

Generally, the purpose of an impounding reservoir is to store a specific waste during periods of low flow in a receiving stream, for subsequent discharge during periods of high flow. The objective of this operation is to maintain a relatively constant concentration of the waste in the receiving stream. Although the total quantity of polluting wastes remains the same, impounding reduces very high concentrations that would occur during periods of low stream flow.

With regard to brine wastes, storage would be accomplished in an artifical reservoir or lagoon. An impervious liner would prevent seepage to underlying ground water. The size of the lagoon is determined based on the historic duration of low flow periods for the specific stream, the daily brine flow, the brine concentration, and the desired chloride level in the receiving stream. The lagoon is provided with a controlling outlet structure to allow discharge proportional to stream flow.

The basic principle of operation of the impounding reservoir is that the chloride concentration in the receiving stream shall be kept as nearly constant as possible. Thus, the discharge from the reservoir must be reduced or eliminated entirely during periods of low flow, while during periods of high flow the discharge must be increased until the concentration of chlorides in the river reaches the design value.

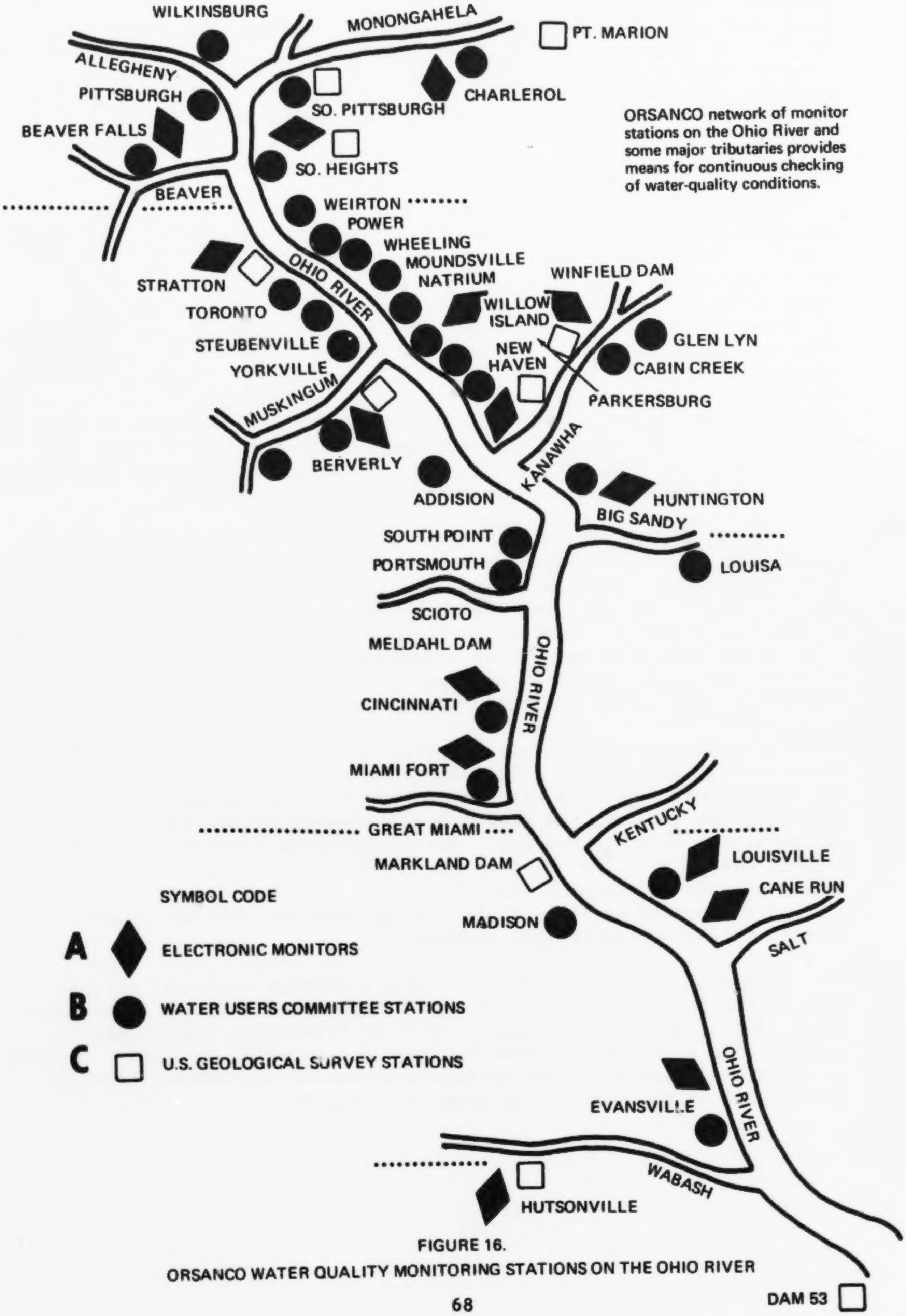
If the flows that will occur during any particular year were known in advance, and if the contribution of each tributary were proportional to its drainage area, it would be possible to operate an impounding reservoir in such a manner as to maintain a constant chloride concentration at any particular point. In practice this is not possible. It is therefore necessary to select the concentration at one point along the river as a basis for operating the reservoir. Using the stream flow records for the chosen point, it is then possible to compute an arbitrary concentration of chlorides that must be maintained at the point. In addition, the system may be automated, if desired, by use of a continuous chloride analyzer installed at the selected point in the stream. An output signal from the analyzer can be utilized to automatically regulate the control device on the reservoir outlet.

In designing a specific system, regulatory agencies must first be consulted for assistance in establishing the chloride level to be maintained in the stream. A minimum historic stream flow is then selected for a specific point in the stream. This value would most likely be based on a minimum monthly average taken for a certain period, such as 10 years. These design parameters, together with information on brine flow and concentration, are utilized to size the impounding basin. A system balance is then developed by use of stream flow records for a two or three year period. If there were no restrictions regarding expenditures or land availability, the system could be designed to maintain a constant year-round chloride concentration in the stream. If restrictions regarding reservoir size do exist, the resulting chloride concentration in the stream at any time can be determined.

The major part of the cost associated with an impounding reservoir is the cost of capital expenditures required for initial installation. Annual operating costs are minimal, particularly where automatic control of discharge is not employed and there are no resulting charges for instrument maintenance. Capital costs include land acquisition, construction of the reservoir and outlet structure with its regulating device, purchase and installation of an impervious liner, and purchase and installation of chloride analyzer, if applicable.

In summary, an impounding reservoir permits the pollution load caused by the disposal of brine wastes in a stream to be spread over the entire year. In this way the average chloride concentration in the stream is not reduced, but peak chloride concentrations during periods of low stream flow are avoided. The improvements created during these periods are balanced by the increase in the chloride concentration during periods of higher stream discharges. Optimum design and operation of an impounding basin would provide constant chloride concentration in the stream during all periods.

An operating installation utilizing a storage reservoir of this type is the Columbia Southern Chemical Corporation Plant at Barberton, Ohio. This program is operated jointly by Pittsburgh Plate Glass Co. and the Ohio River Valley Water Sanitation Commission (ORSANCO). Figure 16 shows the location of the monitor stations of the ORSANCO network. Information on the stations is given in Table 9.



ORSANCO network of monitor stations on the Ohio River and some major tributaries provides means for continuous checking of water-quality conditions.

FIGURE 16.
ORSANCO WATER QUALITY MONITORING STATIONS ON THE OHIO RIVER

TABLE 9

ORSANCO WATER QUALITY MONITOR STATIONS

OHIO RIVER STATIONS

	Mile Point	Type		Mile Point	Type
Pittsburgh, Pa.	2.3	B	New Haven, W. Va.	241.6	A,B,C
South Heights, Pa.	15.8	A,B,C	Addison, Ohio	260.7	B
Stratton, Ohio	55.0	A,C	Huntington, W. Va.	304.2	A,B
Toronto, Ohio	59.1	B	South Point, Ohio	318.0	B
Weirton, W. Va.	62.2	B	Portsmouth, Ohio	350.7	B
Steubenville, Ohio	65.3	B	Meidahl Dam	436.2	C
Power, W. Va.	79.3	B	Cincinnati, Ohio	462.8	A,B
Yorkville, Ohio	83.6	B	Miami Fort, Ohio	490.3	A,B
Wheeling, W. Va.	86.8	B	Markland Dam	531.5	C
Moundsville, W. Va.	111.0	B	Madison, Ind.	559.5	B
Natrium, W. Va.	119.4	B	Louisville, Ky.	600.6	A,B
Willow Island, W. Va.	161.0	A,B	Cane Run, Ky.	616.8	A,C
Parkersburg, W. Va.	183.7	B	Evansville, Ind.	791.5	A,B
			Dam 53	962.6	C

TRIBUTARY STATIONS

	Mile at which Tributary Enters Ohio River	Miles from Sampling Station to Confluence of Tributary with Ohio River	Type
Allegheny River at Kinzua, Pa.	0.0	198.0	C
Allegheny River at Oakmont, Pa.	0.0	12.3	A,B,C
Allegheny River at Wilkinsburg, Pa.	0.0	8.9	B
Monongahela River at Pt. Marion, Pa.	0.0	90.8	C
Monongahela River at Charleroi, Pa.	0.0	42.5	A,B
Monongahela River at South Pittsburgh, Pa.	0.0	4.0	B,C
Beaver River at Beaver Falls, Pa.	25.4	5.3	A,B
Muskingum River at Philo, Ohio	172.2	66.8	B
Muskingum River near Beverly, Ohio	172.2	28.0	A,B,C
New River at Glen Lyn, Va.	-	93.9	B
Kanawha River at Cabin Creek, W. Va.	265.7	72.0	B
Kanawha River at Winfield Dam, W. Va.	265.7	31.1	A,C
Big Sandy River at Louisa, Ky.	317.1	20.3	B
Wabash River near Hutsonville, Ill.	848.0	163.8	A,C

Hauling of Dry Salt or Concentrated Brines

Hauling has been considered as an ultimate waste disposal method that would follow some preconcentration technique (solar evaporation, multistage flash evaporation, multi-effect evaporators, etc.). The costs of preconcentrating to small volumes, such as can conveniently be handled by hauling, are generally so high that the overall combination of disposal schemes becomes unattractive even before the hauling costs are considered. For this reason, and because of the large volumes of brine that would have to be handled at the three selected sites for this study, hauling of dry salts is not economical at any of the sites studied. Instead, abandonment at the evaporation site is used.

The costs for hauling by tank-truck and by rail tank-car have previously been evaluated and have been presented by M.C. Mulbarger. His results in the graphical form (Figure 17) are reprinted here.

Highway tank-truck hauling costs investigations have shown that the method is economical for distances of less than 35 miles and for quantities of less than 17,000 gpd (Koenig, '63).

Rail tank-car hauling is economical for slightly greater distances but is also limited to small quantities (for less than 50 miles conveyance the quantity has to be less than 20,000 gpd).

It can readily be seen that for the distances and quantities investigated in this study, pipelining is the most economical conveyance method, particularly when compared to the combined costs of preconcentrating plus a haul. Hauling is therefore not considered to be a realistic alternative to the other waste disposal methods.

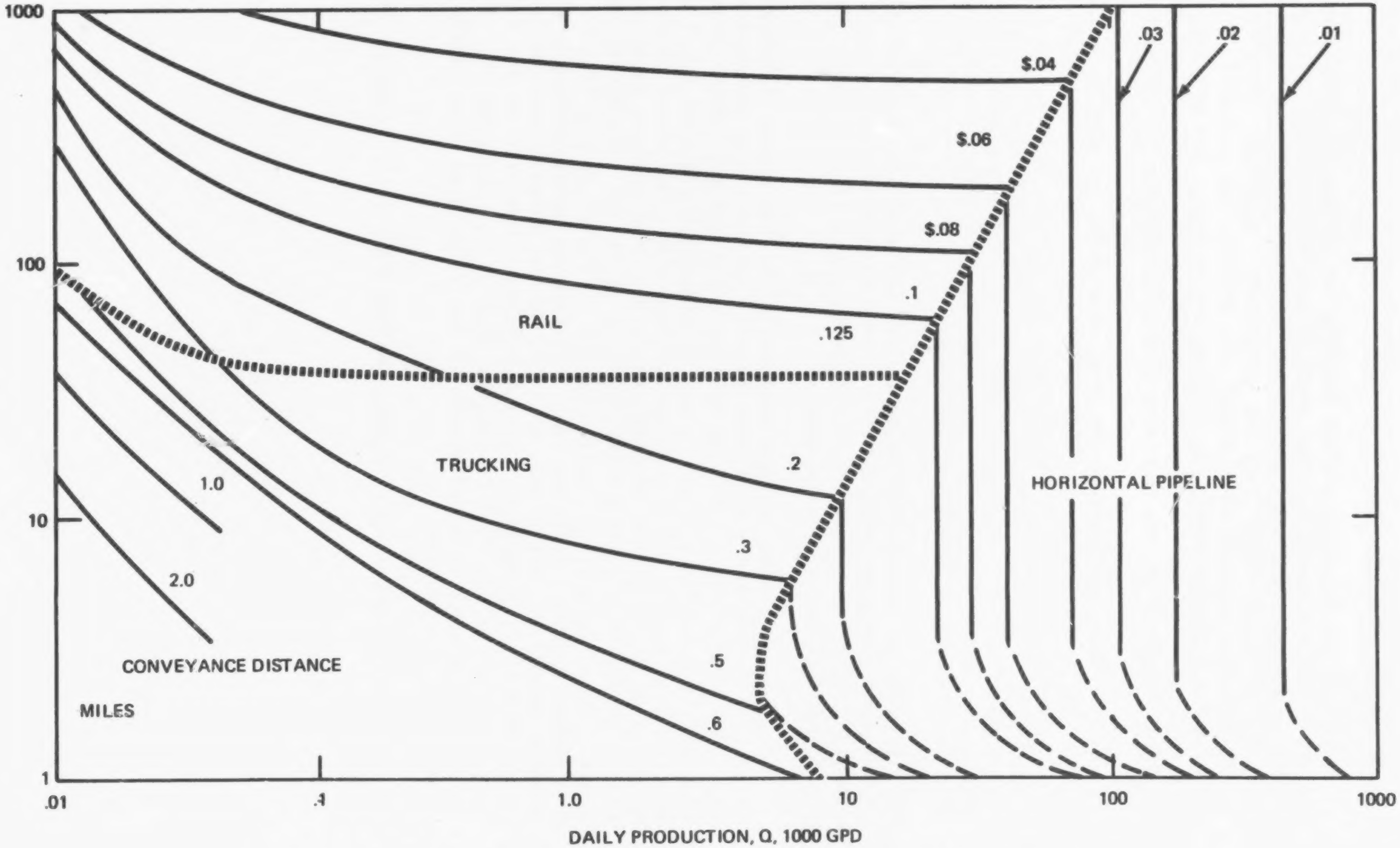


FIGURE 17. COST SURFACE FOR CONVEYANCE: PIPELINE, TRUCK AND RAIL $/1000 gal.-mile)

Brine Desulfation

Purpose and Concept: Brine desulfation is one possible step in the renovation of wastewater, prior to desalinization by electrodialysis or reverse osmosis. The only concern of the study itself is ultimate disposal of the brine waste from the quaternary treatment. However, because the desulfation process will affect the quality and quantity of this ultimate waste, it is necessary to analyze its effect. The motivation for such consideration is the potential advantage to be gained by removal and byproduct recovery of the sulfates initially present in the water.

Desulfation techniques have been considered for use as a pretreatment before desalting seawater by multistage evaporation. Since sulfates cause scale production, the advantages are clear. Higher temperature operation of the evaporators is possible, as is longer duration of operation without shutdown because of the lower rate of scale buildup. Secondly, advantages result from the production of marketable byproducts, such as sulfuric acid and caustic soda.

Similar potential advantages can be analyzed for applicability to wastewater renovation processes. Certain disadvantages may also be present in this case that do not arise in the desalinization of seawater, in addition to known disadvantages.

Conclusions: The extreme toxicity of barium makes it illegal and unfeasible to use the desulfation process in the water renovation system. The other factor which has been considered is the byproduct recovery. Whether the byproduct has any value depends on whether it can compete on the local market. It is estimated that 800 tons/day of H_2SO_4 should be produced to be of any competitive value. The largest water renovation plant of 100 mgd could produce only about 100 tons/day of H_2SO_4 as a byproduct. This is not nearly enough to result in any byproduct credit. The smaller plants of 10 mgd and 1 mgd produce correspondingly smaller amounts of H_2SO_4. Therefore, not only is the desulfation process impractical from the standpoint of barium toxicity, but also it would not even be economical to use on this project were barium not present. Thus, the disadvantages greatly outweigh any of the advantages that can be gained from use of the process.

A Flow Diagram of the Sulfate Removal Process is shown in Figure 18.

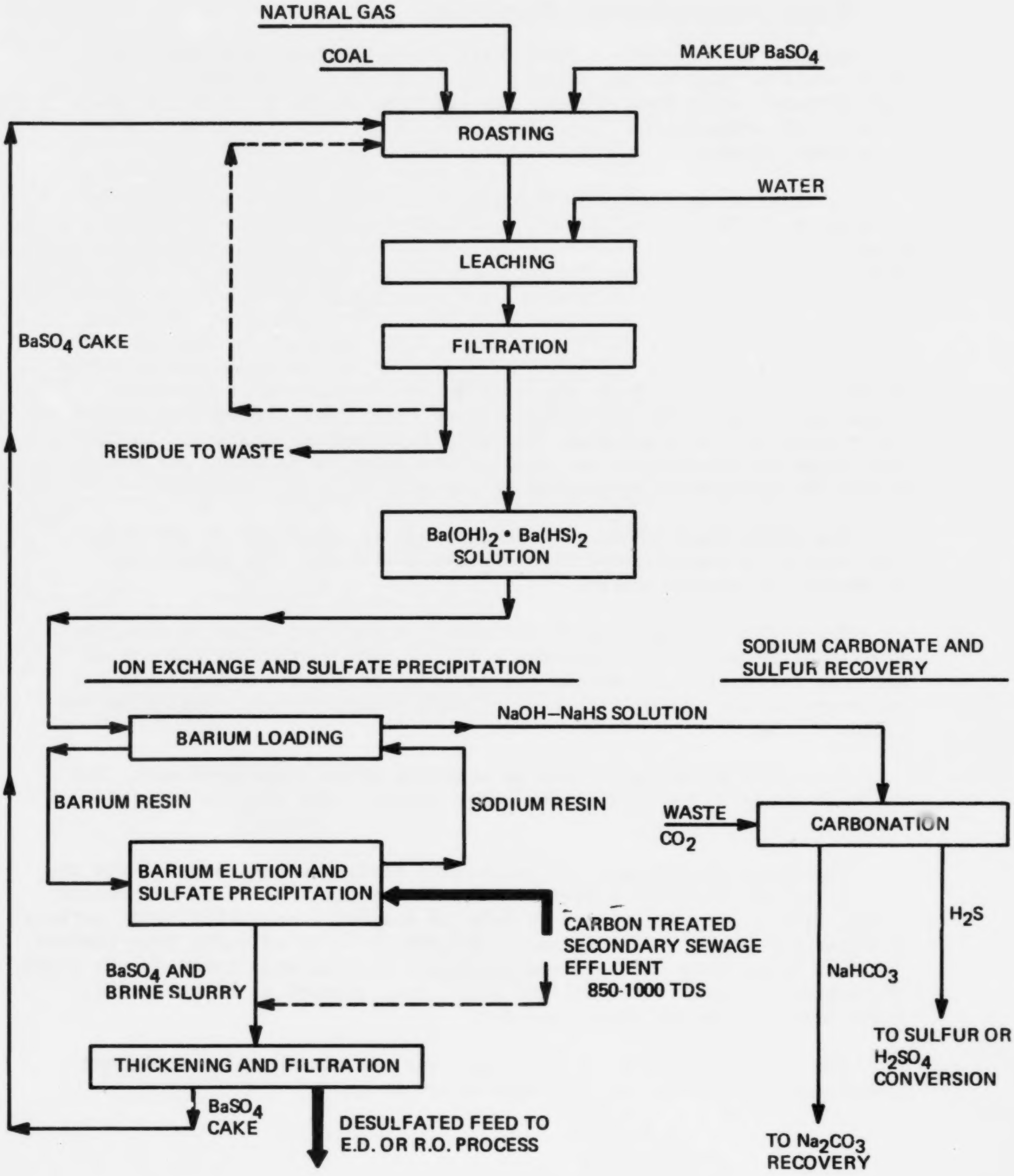

NATURAL GAS
COAL
MAKEUP BaSO4
ROASTING
WATER
LEACHING
FILTRATION
BaSO4 CAKE
RESIDUE TO WASTE
Ba(OH)2 • Ba(HS)2 SOLUTION
ION EXCHANGE AND SULFATE PRECIPITATION
SODIUM CARBONATE AND SULFUR RECOVERY
BARIUM LOADING
NaOH—NaHS SOLUTION
BARIUM RESIN
SODIUM RESIN
WASTE CO2
CARBONATION
BARIUM ELUTION AND SULFATE PRECIPITATION
CARBON TREATED SECONDARY SEWAGE EFFLUENT 850-1000 TDS
BaSO4 AND BRINE SLURRY
NaHCO3
H2S
THICKENING AND FILTRATION
TO SULFUR OR H2SO4 CONVERSION
BaSO4 CAKE
DESULFATED FEED TO E.D. OR R.O. PROCESS
TO Na2CO3 RECOVERY

FIGURE 18. SULFATE REMOVAL PROCESS, FLOW DIAGRAM

Direct Contact Oil-Water Evaporation

Purpose and Concept: A Multistage Flash-Vertical Tube Evaporator plant would be used as the solidifying stage of a waste disposal plant with effluent brine from the tds removal step as the influent or feed brine to the evaporator. The evaporator would concentrate the brine to dryness, forming a filter cake plus distilled (product) water.

In order to produce the proposed filter cake, while avoiding the inherent problems of scaling and corrosion present in an evaporator, a conceptual process utilizing an oil-brine mixture as the influent feed to the evaporator has been investigated. (See Figure 19 on the following page.) In this process, fuel oil and waste brine are mixed and pumped through the condenser section of the evaporator, absorbing the latent heat of the flashing mixture as it proceeds up through the evaporator train. The oil-brine mixture goes to the brine heater where the mixture is heated above the saturation temperature of the first evaporator stage. The oil-brine mixture then flows through the flashing section of the evaporator, becoming the flashing mixture. In the last stage of evaporation the last of the water is removed, and the solids are carried in suspension by the oil.

The water vapor produced in each stage is condensed on the stage tube bundle, preheating the feed oil-brine mixture. The condensate is removed as product water.

The effluent brine-solids mixture from the last stage is fed into an oil-solids separator producing a filter cake containing the solids and some residual oil. Makeup oil is added to the remaining clean oil and recycled for mixing with the influent brine, thereby completing the cycle.

The caked solids are fired in a moving grate steam generator, and the heat produced is used in the brine heater. The slag is carried away for disposal.

Technical Discussion: A feasibility study has been made of the use of oil to heat brine by direct contact in a multistage flash evaporator plant. The plant was assumed to have 16 stages, evaporating equal amounts of brine per stage. Flow rates of 800,000 lb/hr/ft of width were assumed. Oil flow rates into the cold end condenser section were taken as six times the brine flow rate. The oil and brine were assumed to be thoroughly mixed before entering the evaporator.

The required length of two stages was calculated for the highest temperature stage and the low temperature stage.

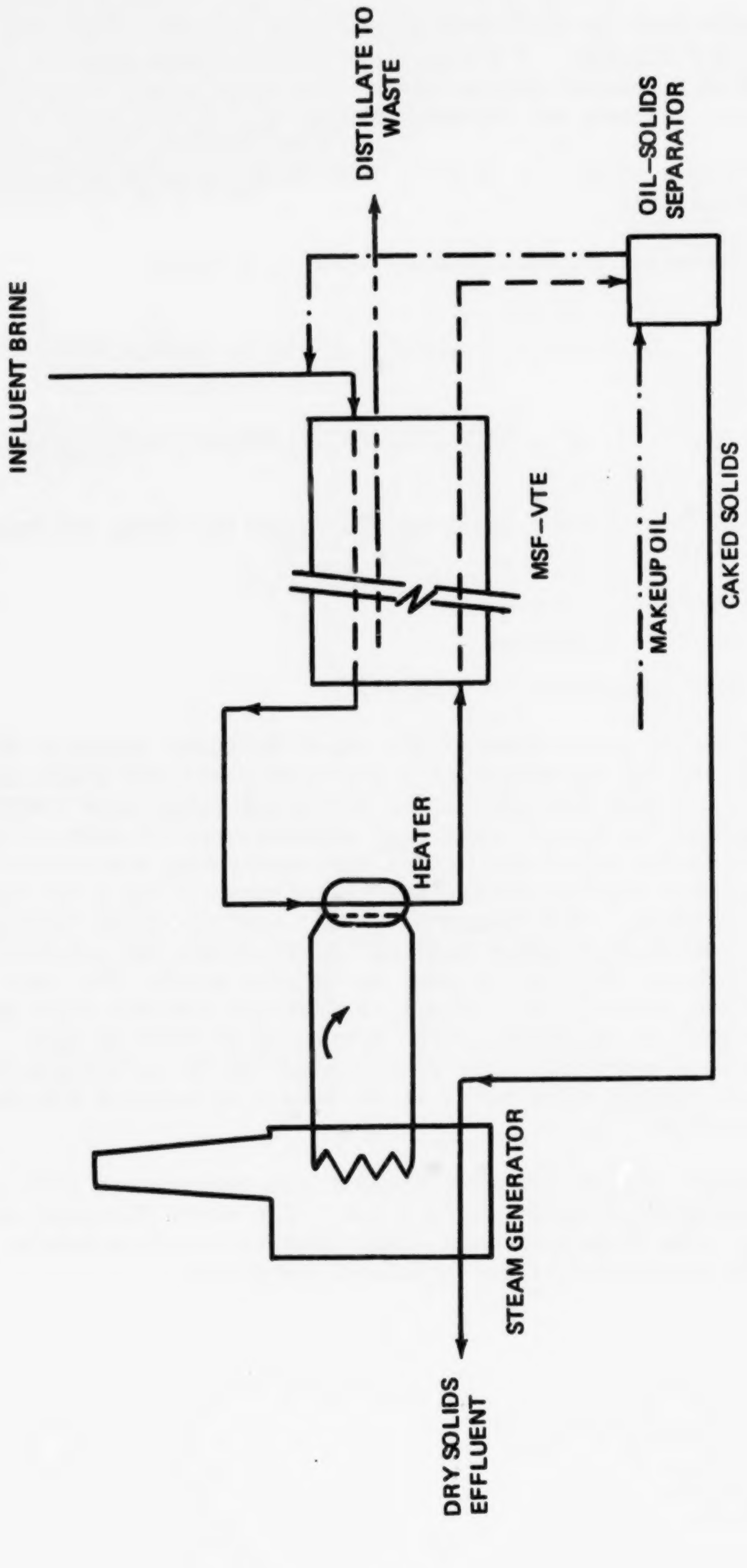


FIGURE 19. DIRECT CONTACT OF OIL–WATER EVAPORATION

Very little data is available on the heat transfer characteristics of direct contact systems. A summary and bibliography applied to the design of a direct contact saline water conversion plant was published by Wilke, Cheng, Ledesma and Porter, 1963.

The advantages of direct contact heat transfer over processes using metallic surfaces are:

No reduction of heat transfer due to scaling

Simplicity of design

Closer temperature approach between the fluids for equivalent heat transfer

On the other hand, the disadvantages of direct contact systems include:

The need to handle separate and circulate large volumes of two fluids

High pumping costs

Large fluid inventory

Product contamination with oil

Results: Using properties of oil-water mixtures computed for a mixture of 6 parts oil by weight to 1 part brine and the usual correlation equations for heat transfer under turbulent conditions through tubes, we computed the inside tube heat transfer coefficient. Since the resistance of the tube wall and of the condensing vapor ($h_o \approx 2000$) is quite small, the inside coefficient calculated is close to the overall clean coefficient. The length of stage required under boiling conditions, assuming reasonable boiling coefficients was calculated as 400 feet which is clearly too long to be practical. The need for packing along the floor of each stage to increase contact area and reduce stage lengths is apparent if the system is to work at all. The paucity of data on evaporation in packed evaporators particularly for multi-component 2-phase flow makes it difficult to predict the amount of packing required.

Conclusions: The use of multistage flash evaporators with direct contact heat transfer between oil and water for waste disposal does not seem feasible. The length of each stage appears to be excessive and the process too questionable for practical purposes.

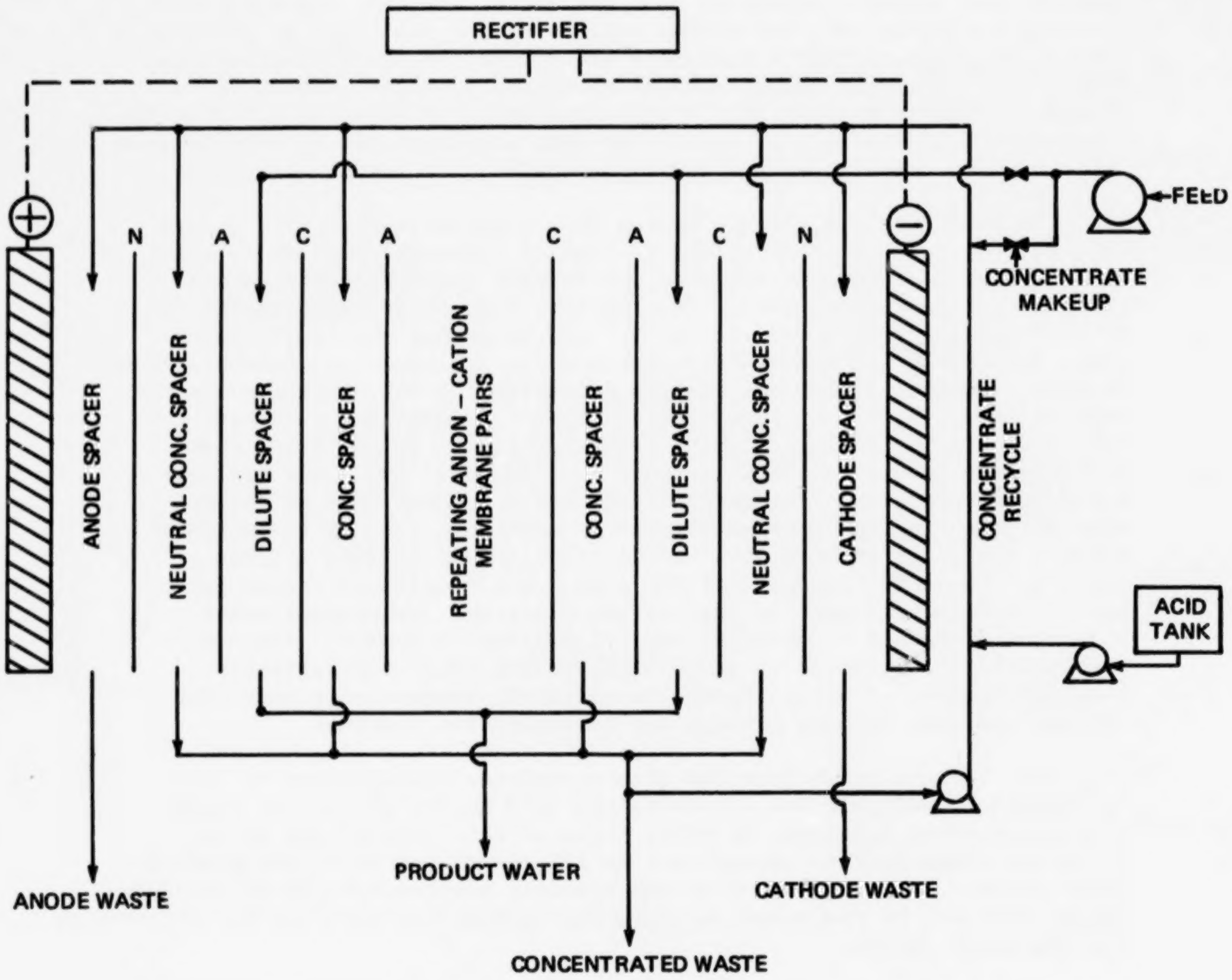
Electrodialysis With Ion-Specific Membranes

This process is one of the methods being used for partially demineralizing water. Basically, the process involves using an induced electric current across a cell containing saline water to separate cationic and anionic components from the water. Cations migrate toward the negative electrode, and anions migrate toward the positive electrode. Cation and anion permeable membranes are placed alternately between the electrodes; ions will concentrate in alternate passages between the membranes and become more dilute in the intervening passages (see Figure 20). Pretreatment of the secondary effluent feed stream to the electrodialyzer would be lime coagulation and softening followed by carbon adsorption.

The pH of the concentrate stream from electrodialysis unit is 4.0. This would have to be neutralized to a pH of 7.0 with soda ash in order to establish a minimum requirement for further post-treatment before being fed into an evaporator. The simplest post-treatment possible is to have the electrodialyzer act as its own softening and conditioning plant for both the concentrate stream and the dialyzate or product stream. However, a modified membrane that is permselective for univalent cations must be used. Seawater concentration plants in Japan use such membranes for concentrating sodium chloride from seawater up to 15% brine that is fed directly to crystallizing evaporators (Yamane, '69). The effluent brine from this kind of electrodialyzer would require only pH adjustment before being fed into an evaporator and could also be concentrated all the way to saturation with sodium chloride and sulfate without scaling. The total hardness of the lime-treated influent stream for the electrodialysis unit is left in the dialyzate, or product water stream, and amounts to about 65 mg/l as calcium carbonate. This minimum hardness is required to render the product water non-aggressive. The concentrate, or brine stream, is softened, because only univalent cations are able to pass through the permselective membrane.

The effluent brine from the electrodialyzer concentrated to 7000 mg/l would be suitable for a non-scaling feed to the multistage flash evaporator plant developed in other parts of this report, and would allow the evaporator to concentrate to 10% tds at the costs for product water shown. In fact, use of an ion-specific electrodialyzer on secondary sewage effluent is equivalent to the basic assumption involved for calculating the evaporator costs.

Alternately, the effluent brine if concentrated to 10% or 15% by the electrodialyzer itself can go directly to ultimate disposal.



C – CATION PERMEABLE MEMBRANE
A – ANION PERMEABLE MEMBRANE
N – NEUTRAL SEMIPERMEABLE MEMBRANE

FIGURE 20. ELECTRODIALYSIS STACK

EL PASO, TEXAS STUDY SITE

WATER SUPPLY AND POTENTIAL NEED FOR REUSE

The city of El Paso is located adjacent to New Mexico and across the Rio Grande from Mexico. It is served by 6 major highways, limited air service and four major railroads with two connecting lines in Mexico.

The elevation at El Paso is about 3750 feet. The El Paso district, which includes all El Paso County, obtains its water from well fields developed in the Hueco Bolson, La Mesa Bolson and in the river alluvium of the Mesilla and lower Mesilla Valley area. The Hueco Bolson extends from the Hueco Mountains on the east to the Franklin Mountains on the west and into New Mexico on the north and Mexico on the south. The La Mesa Bolson of the lower Mesilla Valley is on the west side of the Franklin Mountains (see Figure 21).

The total amount of water that Hueco Bolson and the El Paso River furnish is 46 to 56 million gallons per day in El Paso. The water supply from the Hueco Bolson Basin consists of a lens of fresh water which is floating on salt water. The eastern part of the basin is more saline and is separated from the western part, which contains the fresh water lens, for a distance of 10 to 15 miles.

The city of El Paso is 80% dependent on ground water and 20% on surface water from the river for its municipal supply. The expected future population of El Paso in the years 2000 and 2020 is 615,000 and 913,000, respectively. The Texas Water Development Board has projected future demands for water in El Paso as follows:

Projected Water Requirements to 2020

Total Water Demand (Millions of gallons/day)

	1970	1980	1985	1990	2000	2020
Municipal water	62	78.8	88.7	98.5	123.2	192.5
Industrial water	13	16.5	18.0	19.5	23.13	32.4
Total	75	95.3	106.7	118.0	146.33	224.9

The annual recharge rate into the Hueco Bolson Basin averages 15 million gallons per day; withdrawals total 30 to 35 million gallons a day, including the El Paso River contribution. The water tables are constantly being lowered in the area. To solve this problem, water could eventually be imported from the Mississippi River via the Texas Southwest Water Plan, which is slated for sometime in the future.

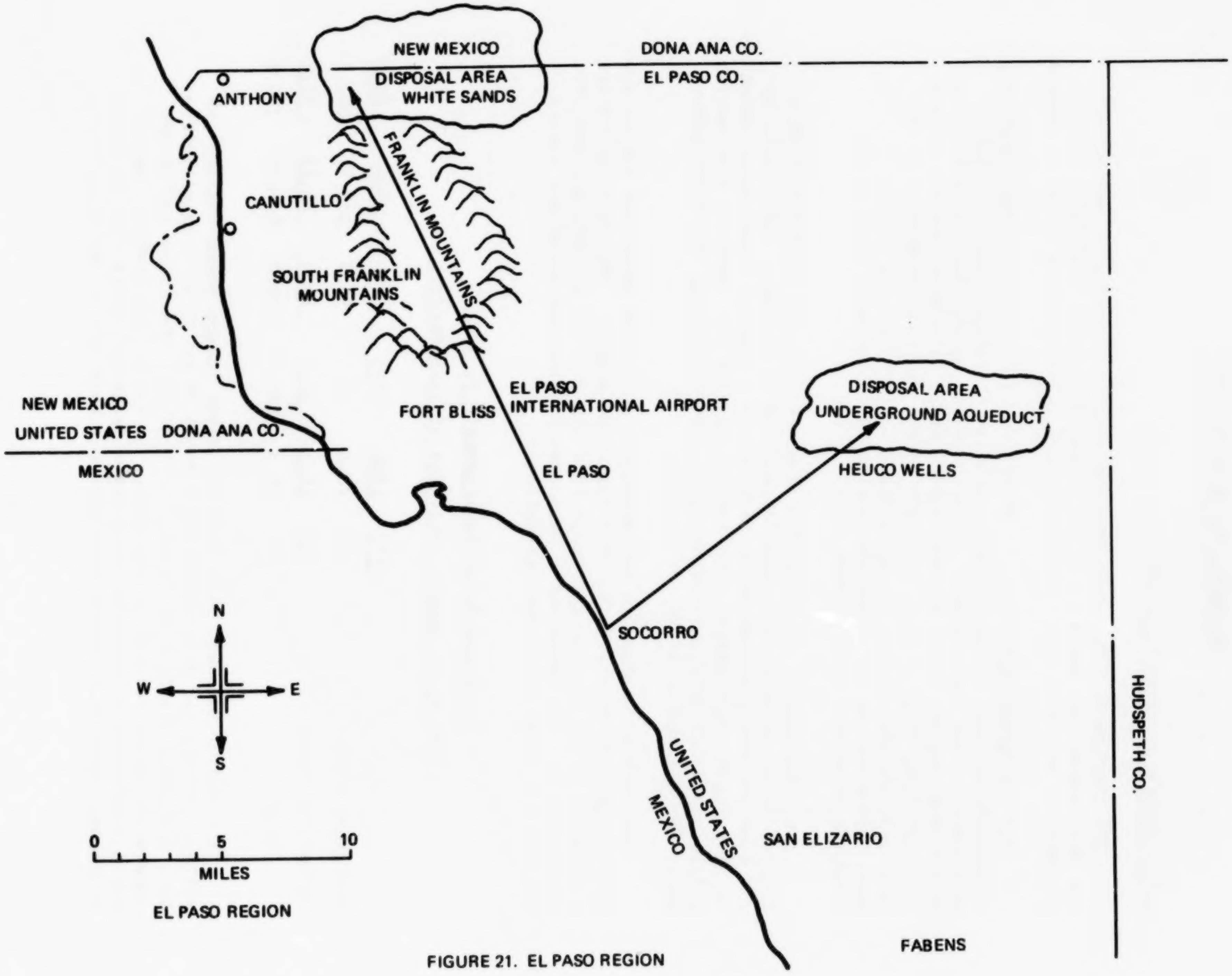


FIGURE 21. EL PASO REGION

The City of El Paso provides its own municipal and industrial water supply. The water production facilities have a capacity of 160 MGD. Maximum daily use is 90 million gallons, and annual use is 16.7 billion gallons. The average per capita daily use is 165 gallons.

CONSIDERATIONS AFFECTING BRINE DISPOSAL

El Paso has two water treatment plants, one for coagulation and filtration of Rio Grande River, the other one for chlorinating ground water.

Sewage treatment is accomplished in three plants utilizing high rate trickling filters to treat the sewage. Total effluent is about 18.2 MGD. Most of the influent to these facilities is from municipal and industrial wastes. There is no municipal reuse of the treated effluent, which ultimately discharges into the Rio Grande.

SPECIFIC BRINE DISPOSAL METHOD (See map of area - Figure 21)

Quaternary Treatment Waste Brines

Assuming that the reclaimed water from the treatment is recirculated and used by the city, this can be combined with all other municipal run-offs under existing water right statutes and returned to the Rio Grande River. By this process no salt beyond that which results from municipal use is added. Under the existing state legislation, this constitutes a legitimate use of water, and so it would be permissible to dump the entire brine effluent from the quaternary treatment into the Rio Grande River at El Paso even though this really evades the pollution problem. At 10 mgd of brine, the tds of the river would, however, exceed 3000 mg/l during the winter months. This would impose undue hardships on downstream users. In addition to the moral or ethical arguments against this disposal method there is another consideration. Laws do change so that what might be legal one day is not legal the next. This is particularly true with laws concerning environmental protection.

More promising alternatives should be considered: Solar evaporation in lined ponds and unlined ponds, disposal by means of deep well at zero well-head pressure, solar evaporation and pipelining combination, and pipelining to a disposal area. For purposes of this study this section presents cost comparisons for the above alternative waste brine disposal methods at capacities of 0.1, 1.0, and 10 MGD of brine, resulting from a tds removal step (quaternary treatment) which concentrates the waste brines to 7000 mg/l tds.

Deep-Well Disposal

Preliminary geological investigations indicate that the salt side of the Hueco-Bolson Basin, 15 miles to the east of the city, could be used as a disposal zone. The eastern part of the basin is more saline and is separate from the western part, which contains the fresh water lens, by a distance of 10 to 15 miles. There is little likelihood at this distance that the injected waste brine would ever contaminate the fresh water supplies in the western side of the basin. Thus, this is a favorable site for deep well injection as an alternative for waste brine disposal. Injection would be made at about 3500 feet. The permeability at that depth is sufficient to allow zero well head pressure to introduce the brine. A profile of the basin is shown in Figure 22.

The cost caluclation for this alternative is based on the following data:

porosity = 30% vol of pore/total vol
height of formation = 100 feet
electrical power cost = 12 mills per kwh
fixed charge rate = 0.05, 0.10
depth of well = 3500 ft
projected life = 30 years
@ 200,000 gpd per well
Brine concentration = 7000 mg/l

From this the deep well injection cost $/Kgal. is:

Capacity MGD	Disposal Cost $/Kgal.		Additional OME Cost $/Kgal.	Total Unit Disposal Cost $/Kgal.	
	FCR=.10	FCR=.05		.10 FCR	.05 FCR
0.1	0.151	.075	0.135	0.282	.210
1.0	0.10	.051	0.065	0.165	.116
10.0	0.106	.054	0.054	0.160	.108

Solar Evaporation

In many arid areas it is possible to evaporate waste brine to dryness in properly lined solar evaporation ponds. El Paso is considered one of the most suitable areas for this process of waste brine disposal. The city owns some 3,000 acres of land which could be used for solar evaporation and, in fact, they are now being used for oxidation ponds. This land was acquired by the city in order to obtain water rights to the surface and ground water for municipal needs. Thus, the water rights of the land itself are expropriated and this land cannot, therefore, be cultivated for farming.

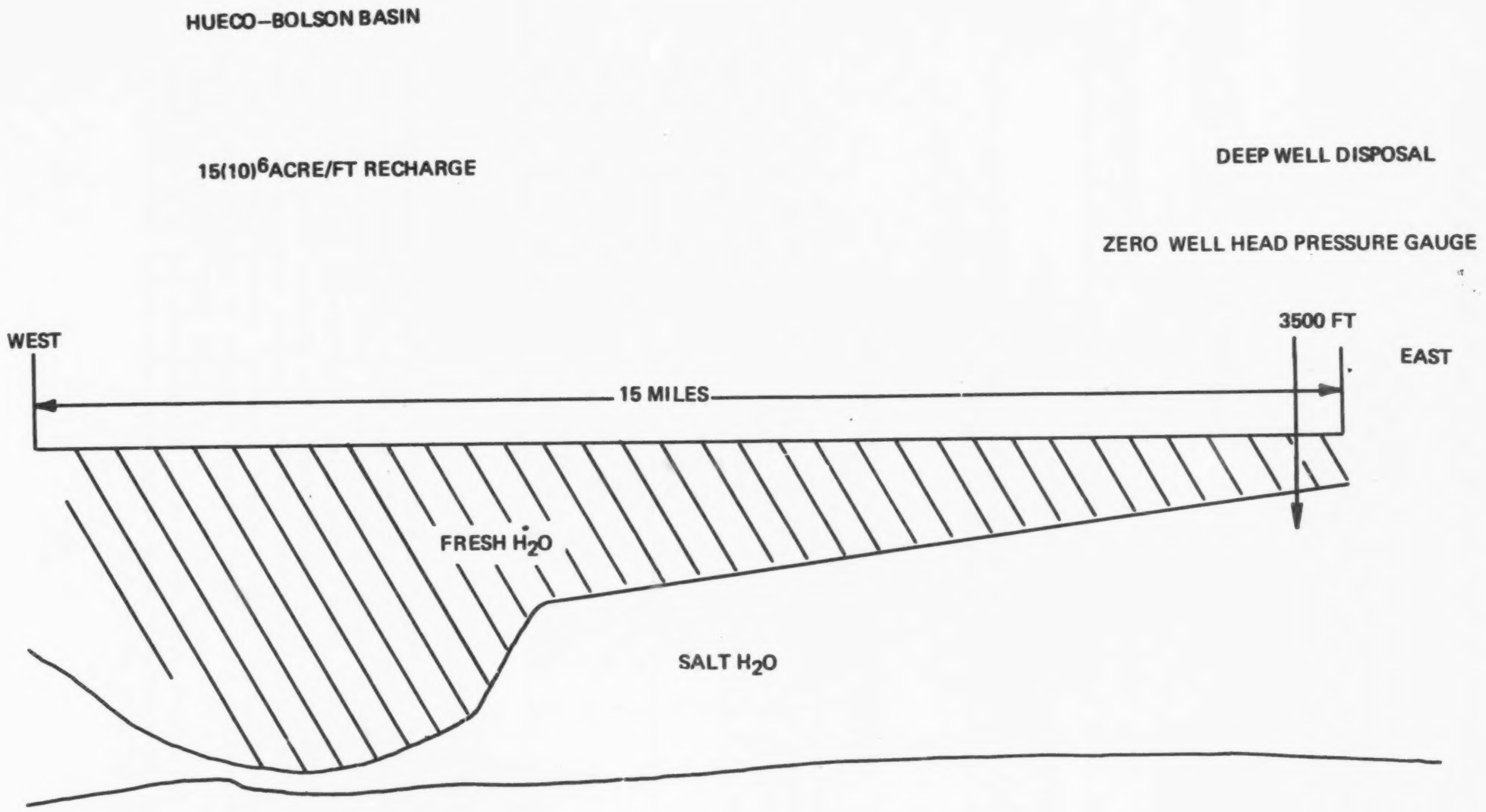


FIGURE 22. DEEP WELL INJECTION AREA OF HUECO–BOLSON BASIN

The costs presented in this process are based upon the assumption of the following:

Waste brine concentration = 7000 mg/l
Total evaporation rate = 105 in./yr
Annual rainfall rate = 12 in./yr
Annual net evaporation rate = 93 in./yr

Lining Cost:

30 mils (PVC) = $0.1322/ft^2 based on 60¢/lb

Land Cost: $250/acre

Dike Cost = $1.00/yd^3 (including equipment, labor and material)

Liner Cover = $0.4/yd^3 (including equipment, labor and material)

Stripping Land Cost = $100/acre

From this the solar evaporation costs $/Kgal. at 5% and 10% fixed charge rates and liner thicknesses are shown in Table 10.

TABLE 10

DISPOSAL COST BY SOLAR EVAPORATION
EL PASO, TEXAS

Brine Quantity, mgd	Fixed Charge Rate, %	Unlined Ponds, $/Kgal.	Lined Ponds (30 mils) $/Kgal.	Lined Ponds (20 mils) $/Kgal.	Lined Ponds (10 mils) $/Kgal.
0.1	10	0.11	0.443	0.35	0.26
	5	0.055	0.22	0.175	0.13
1.0	10	0.047	0.366	0.274	0.184
	5	0.024	0.183	0.137	0.92
10.0	10	0.026	0.342	0.25	0.16
	5	0.013	0.171	0.125	0.08

Pipeline Conveyance

Fifteen miles to the north of the city, the United States Army owns the White Sands proving grounds, both in New Mexico and Texas. This is waste land with no useful aquifers beneath it. Thus, if the Army's permission could be obtained, waste brine from a quaternary treatment could be pipelined over to this area and dumped on the ground, where it would create an artificial salt lake bed. The gross annual evaporation in the area is over 100 inches per year, while the annual rainfall is so light that any salt deposited on the surface would remain on the surface because there is no net run-off from the area. In case the Army will not allow waste brine to be dumped over the area of White Sands, it is still feasible to pipeline waste brine to some other available waste land about 50 miles away from the city of El Paso.

The costs developed for pipeline conveyance disposal alternative, are based on the following data:

Fixed charge rate = 0.05, 0.10

Power cost = 12 mills/kwh

Pumps and accessories cost = $207.5/hp

Installed pipe cost = $10.431/ft
(Basis 12" diameter pipe)

Distance of conveyance = 15, 50 miles

Right of way cost = $5,000/mile

Pipeline conveyance cost in $/Kgal. from El Paso to White Sands at a distance of 15 miles and 50 miles to ultimate disposal site is given in Table 11.

TABLE 11

PIPELINE CONVEYANCE COSTS IN $/KGAL.

Brine Quantity mgd	Pipe Size (In. ID)	Fixed Charge Rate,%	15 Miles		50 Miles	
			Pipeline Conveyance Cost,$/Kgal.	Right of Way Cost $/Kgal.	Pipeline Conveyance Cost,$/Kgal.	Right of Way Cost $/Kgal.
0.1	3	10	0.47	0.205	1.575	0.684
		5	0.27	0.102	0.903	0.342
1.0	8	10	0.19	0.020	0.626	0.068
		5	0.104	0.01	0.346	0.034
10.0	24	10	0.07	0.002	0.24	0.007
		5	0.04	0.001	0.133	0.003

Combined Process - Solar Evaporation and Pipelines

In view of the fact that the accumulation of salt in solar evaporation ponds, lined or unlined, is frowned upon and in some cases prohibited, the best alternative is the combined process of solar evaporation and pipelines. Waste brine would be concentrated in the ponds, thus reducing the volume of waste brine to be transported by pipelining to some waste land for dumping.

Waste brine from a quaternary treatment plant could be pumped from El Paso to available lands about 15 miles from the city. There waste brine would be concentrated in lined solar evaporation ponds and then by pipelining, the waste brine from the pond would be transported to some available waste lands at a distance of about 50 miles for ultimate disposal by dumping.

Costs developed in this alternative are based on the following data:

Cost of pipe (1 ft diameter) = \$10.431/ft
Cost of pump and standby generator = \$207.5/hp
Fixed charge rate = 10% and 5%
Cost of power = 12.0 mils
Cost of right of way = \$5,000/miles
Length of pipeline = 50 miles
Annual net evaporation rate = 93 in./yr
Cost of liner:
  30 mils (PVC) lining = \$0.132/ft^2 based on 60¢/lb
Cost of land = \$250.00/acre

From this the solar evaporation and pipelines combination costs \$/Kgal. for various liner thicknesses are shown in Table 12.

CONCLUSIONS AND RECOMMENDATIONS

Any of the methods for ultimate disposal of waste brine, discussed above, would have to alleviate pollution of natural resources, in order to be acceptable in the area of El Paso. The advantages and disadvantages of these methods depend upon the climate and geology of the disposal area as well as the concentration of the brine. For deep well disposal, pH adjustment is required to render the brine compatible with the soil.

Based upon ultimate disposal of 10 mgd waste brine at 7000 mg/l, using a fixed charge rate of 7%, the following are recommended:

With regard to the total costs, the most economical, but undesirable, brine disposal scheme is to go directly from the quarternary treatment plant into the river. The brine resulting from electrodialysis or reverse osmosis can be combined with all other municipal runoffs under existing water right laws and returned to the Rio Grande River at no disposal cost.

A more desirable economical disposal scheme is to pipeline waste brine to White Sands, 15 miles from El Paso. No net annual runoff and no useful aquifers occur in the area, and so the cost of disposal there by dumping on the ground is simply the pipelining cost. This cost is 5.2¢/Kgal. brine handled, excluding right-of-way.

Deep well disposal of waste brine into the east end of Hueco Bolson basin is also favorable. Waste brine from a quarternary treatment plant will gradually displace, upward and westward, the natural fresh water ground supply in the area. Cost of this scheme is 13¢ per 1000 gallons.

TABLE 12

UNIT COST OF DISPOSAL BY EVAPORATION PONDS AND PIPELINE COMBINATION, EL PASO, TEXAS

EL PASO, TEXAS

Brine Quantity mgd	Fixed Charge Rate %	Lined Ponds (30 mils)		Lined Ponds (20 mils)		Lined Ponds (10 mils)		Additional Cost for Right of Way
		Pipeline Concentration	Disposal Cost,$/Kgal.	Pipeline Concentration	Disposal Cost,$/Kgal.	Pipeline Concentration	Disposal Cost,$/Kgal	
0.1	10	20	0.677	20	0.589	20	0.504	0.685
	5	20	0.353	20	0.309	20	0.266	0.342
1.0	10	20	0.446	20	0.358	20	0.273	0.068
	5	20	0.229	20	0.185	20	0.142	0.034
10.0	10	1.0	0.24	1.0	0.24	20	0.190	0.007
	5	1.0	0.134	1.0	0.134	20	0.097	0.003

Solar evaporation in El Paso is considered one of the most suitable disposal methods. The annual evaporation rate is over 100 inches while the average annual rainfall is light. The City owns some 3000 acres which could be used. Lined ponds would cost between 11¢ and 24¢ per Kgal. of brine evaporated, with liner thickness of 10 to 30 mils.

Other alternatives for waste brine disposal are as follows:

If the area of White Sands is prohibited as a disposal area, it is still feasible to pipeline waste brine to available waste land about 50 miles away from the city of El Paso. This cost is 17.6¢/Kgal. excluding right-of-way.

If the accumulation of salt in lined solar evaporation ponds is prohibited, a combined process of solar evaporation and pipelines would best fit in this case. This cost varies from 13.4¢ to 17.6¢/Kgal. with liner thickness of 10 to 30 mils, excluding right-of-way.

TUCSON, ARIZONA STUDY SITE

WATER SUPPLY AND POTENTIAL NEED FOR REUSE

The City of Tucson is located in Pima County, Arizona between the Tortolita and Tucson Mountains on one side and the Santa Catalinas on the other. The Santa Cruz River flows through the city, entering from the south, and leaving to the northwest. The northwest passage constitutes a subterranean shelf, over which all subterranean waters must flow to leave the valley. Sparse rainfall in the basin constitutes the main source of water at present, since the river is usually a dry bed. The city takes its water supply from wells in the valley. Some recharge of the underground supply occurs as a result of crop irrigation, utilizing the municipal wastewater, but this is insufficient to stabilize the water table, which has been declining for many years. Reuse of available supplies by irrigation is almost 100%.

To offset an expected substantial population growth in the near future, the city expects to import some 300 mgd of Colorado River water from Parker Dam by the year 2,000, which together with anticipated reuse and recharge will stabilize the water table at that time. The Colorado River water by then is expected to contain 1,000 mg/l tds, which, together with an expected 300 mg/l tds added by a single city use must be removed from the valley, either by irrigation return flow to the already contaminated Gila River, or by desalination in order to prevent a major buildup of salt in the ground water supplies.

CONSIDERATIONS AFFECTING BRINE DISPOSAL

Tucson has a sewage plant effluent capacity of 36 mgd. The influent to this facility is primarily municipal waste with very little industrial waste, which has picked up only 175 mg/l tds as a result of the city use. The effluent, after primary and secondary treatment, is used to irrigate cotton and wheat crops in the valley. A small amount of irrigation runoff, together with any bypassed sewage effluent enters the Santa Cruz River bed, ultimately reaching and drying up in the Gila River. Most of the irrigation water, however, percolates underground to serve as recharge for the reservoir. As a consequence, the nitrogen content of the ground water has been increasing recently. To prevent this buildup, the city is installing an anaerobic denitrification facility as a tertiary sewage treatment step. Salt buildup in the ground water supplies, however can be expected to become an increasingly aggravated problem, until a means for removing it can be found. Only desalination of the wastewater, with ultimate disposal of the resultant brine offers a permanent solution.

SPECIFIC BRINE DISPOSAL METHODS (See map of area - Figure 23)

Injection Wells

The Tucson basin is composed of sandy alluvium, the depth of which has never been measured. One well, drilled to 2,000 feet, has produced fresh water. The permeability of the soil is excellent. Beneath the useful strata, it is surmised that salt water exists, probably at about 3,000 feet depth. Injection of brine at about 3,500 feet depth would not be expected to create deleterious effects for a relatively long period of time. It is thought that any salt added to the aquifer would remain at the bottom, and would serve to raise the water table of useful water, which would float as a lens over a basin of salt water.

The costs for this disposal method are based on the following data:

Depth of well = 3500 feet; height of formation 10 feet
Porosity, P = 30% (void volume/total volume)
200,000 gpd/well; electrical power cost 12 mills/kwhr
Project life = 30 years; fixed charge rates .05 and .10

Capacity MGD	Disposal Cost $/Kgal.		Additional Oper. & Maint. Cost	Total Cost $/Kgal.	
	10% FCR	5% FCR		5% FCR	10% FCR
0.1	0.1506	.075	0.1350	0.210	0.2856
1.0	0.0996	.051	0.0652	0.116	0.1648
10.0	0.1062	.054	0.0540	0.108	0.1602

Pipelining

Fifty miles to the East of Tucson is a closed salt basin, known as the Wilcox Playa. In the middle of this basin is a sulfur lake formerly used as a health resort.

It is feasible to pipeline waste brine to this area. This would not require soil preparation or liners, since the area cannot be further contaminated by additional salts.

The net effect of accumulating salt in this area would be to create a new great salt lake basin, which might ultimately be mined for salable salts.

The costs that have been developed for this disposal method, were based on the following data:

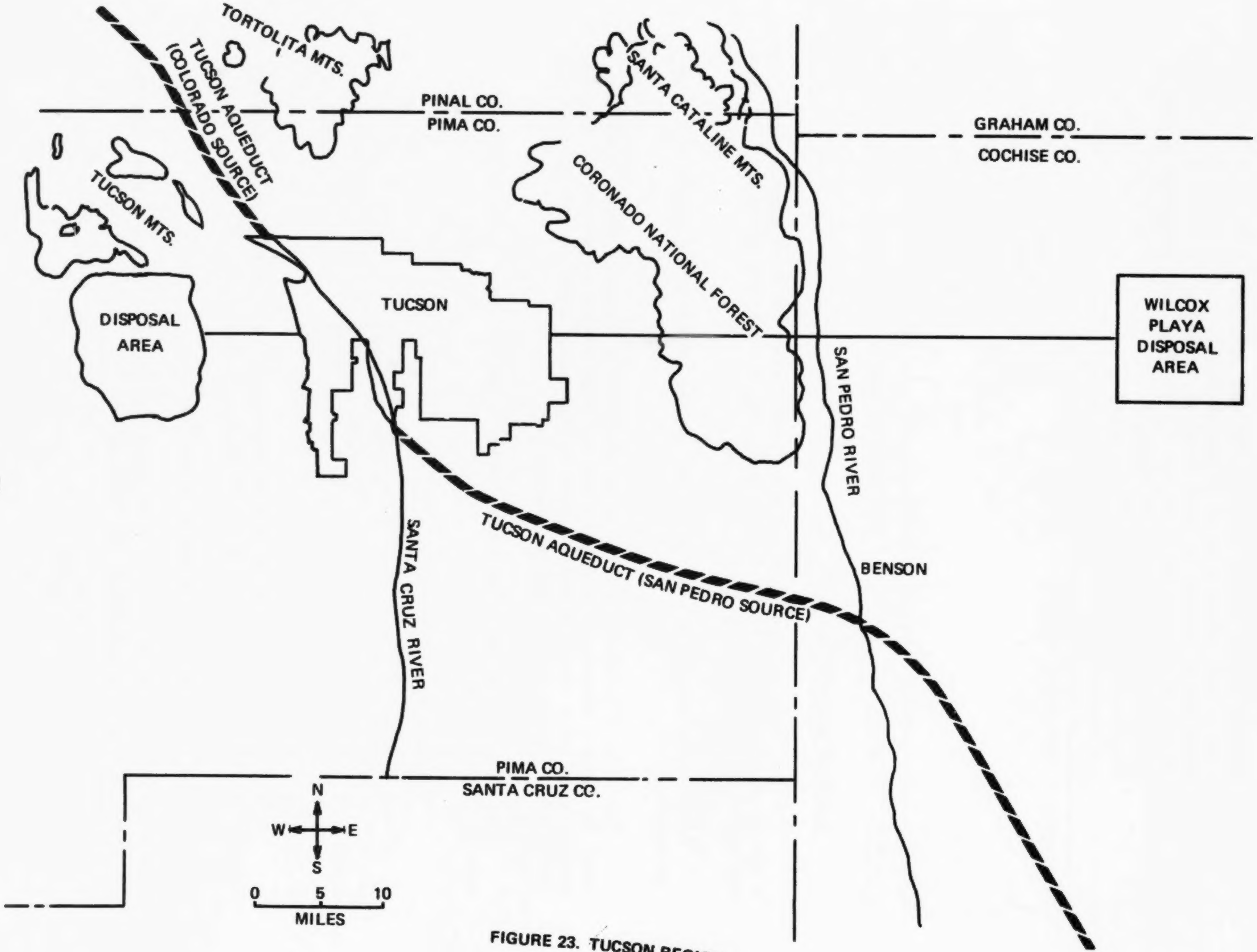


FIGURE 23. TUCSON REGION

Annual fixed charge rate = .05 and .10
Electrical power cost = 12 (mills/kwh)
Pump and accessories cost = 207.5 ($/hp)
Installed pipe cost = 10.431 ($/ft (for 12" diameter pipe))
Conveyance distance = 50 (miles)
Right-of-way cost = 5000 ($/mile)

Capacity MGD	Optimized Pipe Size, In.	Conveyance Cost,$/Kgal.		Right-of-Way Cost,$/Kgal.	Total Cost,$/Kgal.	
		10% FCR	5% FCR		5% FCR	10% FCR
0.1	3	1.576	.903	0.684	1.587	2.260
1.0	8	0.626	.346	0.068	0.418	0.694
10.1	24	0.240	.133	0.007	0.140	0.247

Solar Evaporation Ponds

The net evaporation rate in the Tucson area is very high. Based on a gross evaporation rate of 95 in./year and a rainfall of only 5 in./year, the net evaporation rate is 90 in./year.

With such high evaporation rates, solar evaporation ponds are very attractive.

There is limited land available within the valley, which could be used for solar evaporation ponds, but linings would be mandatory so that accumulated salt left in these after abandonment could never contaminate the useful ground water supply beneath.

The costs developed for this disposal method have been based on the following data:

Net evaporation rate = 90 (in./year)
Annual fixed charge rate = .10 and .05
Cost of land = 250 ($/acre)
Dike fill cost = 1.00 ($/yd^3)
Liner cover cost = 0.40 ($/yd^3)
Stripping land cost = 100 ($/acre)

PVC liner costs based on 60¢/lb have been developed for three different thicknesses of the liner - 10 mils, 20 mils, and 30 mils.

30 mils (PVC) lining = $0.132/ft^2

The costs have also been developed for unlined solar ponds, which, however, are not recommended for this site.

For aesthetic as well as economic reasons salt and concentrated brine cannot be allowed to accumulate indefinitely at solar pond locations within the Tucson Valley itself. These would have to be hauled or pipelined away to a suitable non-leaching sump, such as the Wilcox Playa, 50 miles to the East. Therefore, at Tucson, solar evaporation must be considered as a pre-concentrating technique preceding ultimate disposal and not as a complete disposal method in itself. Either the costs of a dry haul or of a pipeline must be added to the solar pond costs to obtain the complete costs of ultimate disposal.

The solar evaporation costs \$/Kgal. at 5% and 10% fixed charge rate and liner thickness are shown in Table 13.

TABLE 13

UNIT DISPOSAL COST BY SOLAR EVAPORATION, TUCSON, ARIZONA

Brine Quantity mgd	Fixed Charge Rate,%	Unlined Ponds \$/Kgal.	Lined Ponds (30 Mils) \$/Kgal.	Lined Ponds (20 Mils) \$/Kgal.	Lined Ponds (10 Mils) \$/Kgal.
0.1	10	0.11	0.455	0.36	0.267
	5	0.055	0.228	0.18	0.133
1.0	10	0.048	0.378	0.283	0.19
	5	0.024	0.189	0.141	0.095
10.0	10	0.027	0.354	0.258	0.165
	5	0.014	0.177	0.13	0.083

Solar Pond and Pipeline Combination

The immediate problem facing the city of Tucson is nitrogen buildup in the ground water supplies, due to irrigation with secondary sewage effluent. Preliminary results of experiments with algae ponds in the area indicate that in these ponds the pH of the effluent builds up to 8.2, and that at this pH, considerable nitrogen is released into the air in the form of ammonia. Conceivably, the residual nitrogen which is in the form of nitrates could next be removed by anaerobic denitrification, down to an acceptable level which could be utilized by the roots of the plants under irrigation. More extended tests of algae ponds in other areas, however, have demonstrated that continuous removal of ammonia nitrogen is possible if, and only if, the algae bloom is continuously harvested; otherwise, decay of the accumulating algae will produce a serious problem. The overall costs of the operation usually turn out to be quite high.

Quaternary treatment of the sewage effluent so as to remove plant nutrients offers a way out of this quandary. The quaternary treatment considered under the ground rules of this study is electrodialysis, preceded by lime clarification and carbon adsorption. A study of the composition of the sewage effluent from the city of Tucson, made herein in connection with ion-exchange pretreatment for Multistage Flash Evaporation, has shown that lime clarification is sufficient to remove the phosphate nutrients. Following this step, the electrodialysis process can be arranged stagewise, using permselective membranes that are specific for monovalent ions (Yamane '69), such as ammonium and nitrate ions, so that these ions, instead of winding up in the renovated wastewater, will be concentrated in the brine stream. As the brine stream is to be pumped out of the area and into the Wilcox Playa Basin, this procedure allows a portion of the nitrogen pollution problem to be solved at no additional cost, along with the main salt pollution problem.

There is an additional saving to be made over pipelining away the waste brine from the electrodialysis unit in the "as received" condition. The best operation of the electrodialyzer will result in a brine that has been concentrated only about 10 times, to about 7,000 mg/l. This can be preconcentrated before pipelining to any desired degree by the use of local lined solar evaporation ponds. The computer program which has been developed for this purpose is useful to determine for the local site conditions whether preconcentration is economical.

For the three design capacities of this study, the costs developed for the combined solar preconcentrating and pipeline disposal method by means of the computer are shown in Table 14 and have been based upon the same data as the solar pond and separate pipeline disposal methods.

TABLE 14

UNIT COST OF DISPOSAL BY EVAPORATION PONDS AND PIPELINE COMBINATION, TUCSON, ARIZONA

Brine Quantity mgd	Fixed Charge Rate,%	Lined Ponds (30 Mils)		Lined Ponds (20 Mils)		Lined Ponds (10 Mils)		Additional Cost for Right-of-Way $/Kgal.
		Pipeline Concentration	Disposal Cost,$/Kgal.	Pipeline Concentration	Disposal Cost,$/Kgal.	Pipeline Concentration	Disposal Cost,$/Kgal.	
0.1	10	20	0.689	20	0.599	20	0.510	0.685
	5	20	0.359	20	0.314	20	0.27	0.342
1.0	10	20	0.457	20	0.367	20	0.278	0.068
	5	20	0.234	20	0.189	20	0.145	0.034
10.0	10	1.0	0.24	1.0	0.24	20	0.195	0.007
	5	1.0	0.134	1.0	0.134	20	0.100	0.003

CONCLUSIONS AND RECOMMENDATIONS

Based upon ultimate disposal of 10 mgd waste brine at 7000 mg/l, using a fixed charge rate of 7%, the following are recommended:

As a temporary measure until imported Colorado River water from Parker Dam is available, deep well disposal of waste brine into the Tucson basin at 3500 ft would cost 13¢/Kgal.

Another economical disposal scheme is to pipeline waste brine into a salt lake at Wilcox Playa, fifty miles to the East of Tucson. This cost is 18¢/Kgal. brine, excluding right-of-way.

If sufficient land can be found near Tucson, lined solar evaporation ponds with salt abandonment at the site are feasible. Ponds would cost between 11.6¢ to 25¢ per Kgal. with liner thickness of 10 to 30 mils.

To provide salt disposal equal to the salt entering rate expected by 1985, so that the imported water from the Colorado River will not irreversibly contaminate the ground water supplies, the most economical alternative is using a combination of local solar ponds with a pipeline to convey the residual brine to the Wilcox Playa, 50 miles to the East. This cost varies from 14¢ to 18¢/Kgal. depending upon liner thickness of 10 to 30 mils.

DENVER, COLORADO STUDY SITE

WATER SUPPLY AND POTENTIAL NEED FOR REUSE

The City of Denver is located in Jefferson County, Colorado, some 100 miles south of Wyoming and approximately 100 miles north of Pueblo. Interstate highways 25, 49, 70, 80, 85, 87, 287 and 285 converge in Denver.

With regard to the water supply, the largest contribution to Denver's available water is the Blue River Diversion System. The system comprises the huge Dillon Dam and Reservoir with a capacity of 1/4 million acre feet annually, almost equal to that of all Denver's other reservoirs put together, and the 23 mile long Harold D. Roberts Tunnel under the Continental Divide, the world's longest underground tunnel. Water first came to Denver through the Roberts Tunnel on July 17, 1964. In 1968 this and other water supplies yielded more than 190 mgd of water, with 15 mgd being added to storage.

The transmountain storage facilities furnish over half the water used in Denver. The metropolitan area itself has small water utilities utilizing shallow wells and deep well pumping. Surface runoff is the main supply. Agricultural ditches, which use the majority of the waters, furnish drinking water by appropriation. The South Platte River itself is presently overappropriated.

Projections of water requirements for Denver yield an estimate of municipal and industrial water for 2008 about three times as great as for 1969. In order to provide for these needs, it has been recognized that desalting should be considered as one of the possible sources of future supply. However, according to the Denver Board of Water Commissioners, studies indicate that new water supplies primarily from the Colorado River System will be available. Coupled with transmountain water, a small additional supply through purchase of water rights should adequately supply the projected municipal and industrial needs to the year 2008. Projection of future demands for water in Denver is as follows:

Projected Water Requirements to 2008

Total Water Demand (MGD)

	1970	1980	1985	1990	1995	2008
Treated Water	167	212.6	236.8	289.2	355.4	473.7
Raw Water	12.8	14.4	15.9	16.3	16.8	19.8
Subtotal	179.8	227	252.7	305.5	372.2	493.5
Subtotal + 7.5% Operating Loss	193.3	244	271.6	328.4	400.1	530.5

The current and expected future growth trend of the population of Denver is as follows:

Current and Projected Population to 2008

1968	1988	2008
1,043,200	1,719,000	2,404,300

With regard to rates charged for water, there are presently 90,000 flat rate residences that were using water prior to metering. The balance of the residences are now metered.

CONSIDERATIONS AFFECTING BRINE DISPOSAL

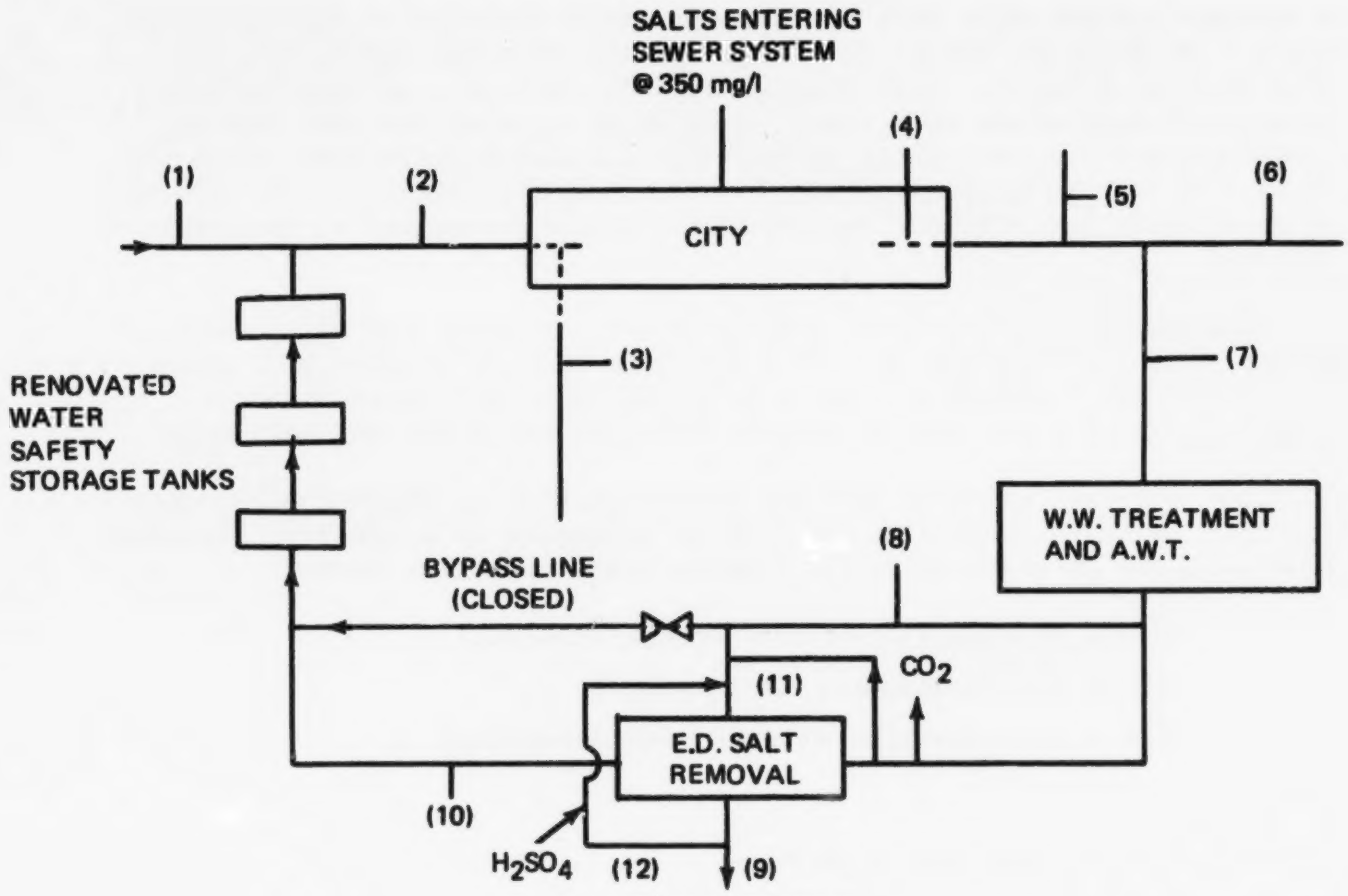
The City and State laws forbid the reuse of water of the South Platte River because the interest of the people has always been in pristine quality water. The existing legislation, however, did not anticipate transmountain water, furnished through the aqueduct from the Colorado River supply near its source, the Blue River. Thus, the only water that can be considered for reuse is Colorado River water. Even the reuse of Colorado River water may require court interpretation, since existing legislation is ambiguous.

The South Platte River water and ground water contain 150 mg/liter tds average, while for the diverted water from the Colorado River there are only 50 mg/l tds, a negligible amount. The consumption in 1968 was 148 mgd, of which half came from the Colorado River. Thus, about 75 mgd would be the amount that could be utilized in a direct water reuse scheme, if the reuse of Colorado River water is determined to be lawful. Projected to 197 mgd reuse for the future, this would involve a disposal scheme for 10 mgd of wastewater containing about 7000 mg/l tds from the reuse, provided salt removal by electrodialysis or reverse osmosis is used. This would require, in turn, a makeup from the Colorado River of only 10 mgd.

A smaller wastewater renovation plant of 100 mgd capacity has already been proposed for Denver. Under a study being pursued by the University of Colorado, such a plant could be located at the North Metro Filter Station, 104th Avenue near the South Platte River, where acreage already owned by the City could be utilized.

Hypothetical 197 MGD Water and Salt Balance for Denver, Colorado based on Renovating a Portion of its Wastewater for Reuse

The schematic diagram, Figure 24, illustrates a reasonable concept of water renovation. It covers most of the situations which would be met under a number of different circumstances. The bypass line would be needed in a situation where not much salt has to be removed from the renovated water. It is not practical to run all of the water through



MATERIAL STREAMS

	DESCRIPTION	STREAM VOLUME, MGD	SALT CONCENTRATION mg/l
1.	PRISTINE WATER INPUT	$W = 10$	$C_w = 115$
2.	TOTAL INPUT TO CITY	$Q = 197$	$C_i = 500$
3.	INPUT WATER WHICH DOES NOT ENTER SEWERS	$y = 0$	$C_i = 500$
4.	TOTAL SEWAGE LEAVING CITY	$q-y = 197$	$C_o = 850$
5.	STORM WATER AND INFILTRATION	$s = 0$	$C_s = -$
6.	WASTEWATER LOST OR NOT RECYCLED	$p = 0$	$C_p = -$
7.	WASTEWATER FED TO W.W. TREATMENT AND AWT	$q-y+s-p = 197$	$C_p = 850$
8.	BYPASS OF SALT REMOVAL	$n = 0$	$C_p = -$
9.	BRINE DISCHARGED FROM SALT REMOVAL UNIT	$b = 10$	$C_b = 7000$
10.	RENOVATED WATER RECYCLE	$r\ (=q-y+s-p-n-b) = 187$	$C_r = 521$
11.	CONCENTRATE STREAM FEED	$F = 10$	$C_f = 850$
12.	CONCENTRATE STREAM RECYCLE	$C = 177$	$C_b = 7000$

FIGURE 24. SCHEMATIC REPRESENTATION OF WATER AND SALT BALANCE FOR 197 MGD WASTE WATER RENOVATION PLANT DENVER, COLORADO

a reverse osmosis unit just to remove a small fraction of the dissolved salts. It would be better to remove a large fraction and blend back with renovated water. With electrodialysis, however, as used in this study, a single stage will remove only up to half of the tds, and so circulating all of the water through the electrodialysis unit turns out to be the most economical method for influent concentrations of 1000 mg/l or less, with the effluent concentration being controlled by the applied voltage.

The addition of sulfuric acid to avoid carbonate scaling is necessary with electrodialysis. This normally adds an increment of about 50 mg/l in total dissolved solids as a result of the replacement of carbonate by sulfate. No correction for this increment has been made.

The material balances for the representation as shown involve many variables which cannot be fixed without reference to a specific location. Consideration is given to the following simple case for Denver:

$S = 0$ (transpiration loss equals rainfall)

$P = 0$ (all wastewater is recycled)

$n = 0$ (no renovated water bypass desalting)

$y = 0$

Material balances may now be made.

Overall Water Balance:

$$W = y + b \tag{1}$$

Overall Salt Balance:

$$W\,C_w + (q-y)\,(C_o - C_i) = y\,C_i + b\,C_b \tag{2}$$

Dividing both equations through by "b", we obtain

$$\frac{W}{b} = \frac{y}{b} + 1 \tag{3}$$

$$\frac{W}{b}\,C_w + \left(\frac{q-y}{b}\right)(C_o - C_i) = \frac{y}{b}\,C_i + C_b \tag{4}$$

Eliminating $\frac{y}{b}$ from equation 4 gives

$$\frac{q-y}{b} = \frac{C_b - C_i}{C_o - C_i} + \left(\frac{W}{b}\right)\left(\frac{C_i - C_w}{C_o - C_i}\right) \tag{5}$$

A 197 mgd plant for waste renovation with tds removal, if installed, would increase the total water supply to 335 million gpd, which would be sufficient for the year 1990 without requiring any greater withdrawals from the Colorado River than at present. Denver's plans to use larger amounts than this in the future from the Blue River Diversion System are not popular in the western parts of the state.

The existing facilities have a treated water capacity of 287 mgd as follows:

Moffatt	63	
South Platte	72.7	
Blue River	151.3	(tributary of Colorado River)
	287 mgd	

The per capita consumption is steadily increasing primarily as a result of the rising standard of living. The present 224 gpd per capita is estimated to reach 255 gpd in the year 2000.

The use of transmountain Colorado River water in Denver area, however, will not be without difficulties in the future. The people in western Colorado are already objecting to this exportation because the potential production of petroleum oil from oil shale is becoming of increasing importance to the state's economy in the west. This production requires forty to fifty barrels of water per barrel of oil produced for the purposes of extracting the oil from the shale. During this use, a small portion of water is used for cooling, but the largest amount is used in direct contact with the hot shale which contains 25 to 60 percent of oil by weight. The resultant water is highly contaminated with soluble material from the shale, notably sodium chloride and bicarbonate. Thus, when oil extraction from this shale becomes operational in western Colorado, water supply and salt pollution problems will arise there, and means must be found for augmenting rather than diminishing the available supply of Colorado River water in western Colorado.

SPECIFIC BRINE DISPOSAL METHODS (See map of area - Figure 25)

Of the methods for ultimate disposal of brine, presuming that permission can be obtained for reuse of water in the Denver area, the lined solar evaporation pond and the combined process of forced and solar evaporation seem to be the only methods that would be acceptable to every agency concerned. The use of deep wells for disposal, because of the earthquake problem and of the direct correlation of frequency and intensity of earthquakes with the rate of injection by the Denver Arsenal, forecloses completely the possibility of well injection into the fractured precambrian breccia stratum. Higher strata all communicate directly with ground water in low lying areas outside of Denver and hence would not be suitable. A future possibility for limited storage of very concentrated brines would be hydrofracturing of some impermeable stratum

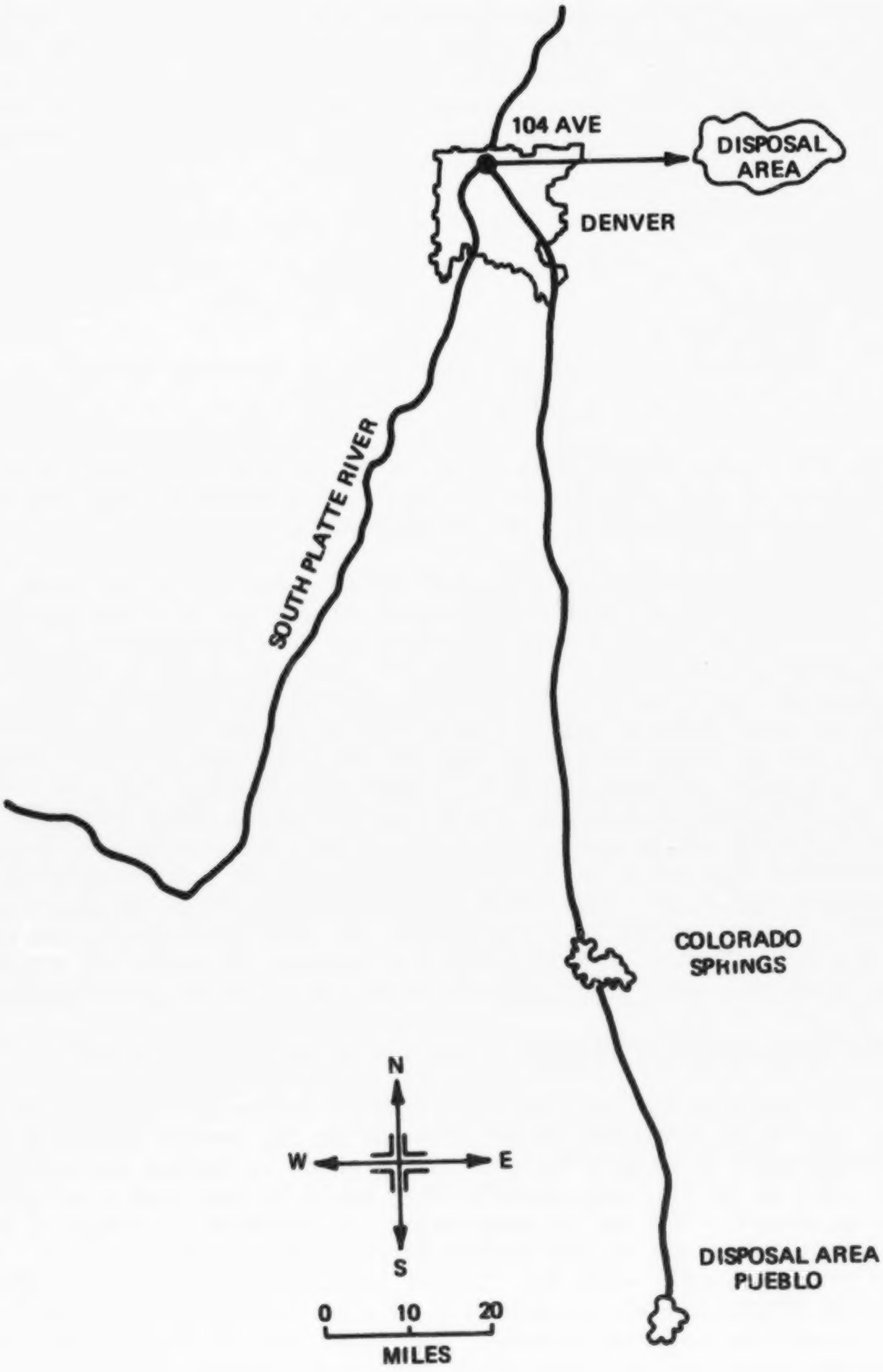


FIGURE 25. DENVER REGION

outside the earthquake zone and sealing the brine permanently into the cavity thus formed. Pipelining of brines to the ocean or the great salt lakes is, of course, impossible. For purposes of this study, the following cost comparisons for acceptable alternative waste brine disposal methods at a capacity of 0.1, 1.0, and 10 MGD of brine are presented:

Solar Evaporation

The net annual evaporation in the vicinity of Denver is about 29 inches, while the average annual rainfall is 13 inches. In the Pueblo area, some 100 miles south of Denver, the net annual evaporation rate is much higher, 36 inches, while the annual rainfall is about the same, 12 inches. Thus, the alternatives compare solar evaporation costs in lined ponds located immediately east of Denver, with costs of a pipeline to Pueblo and solar evaporation east of Pueblo in a desert area.

The costs presented for these schemes are based upon the following assumptions:

Waste brine concentration = 7,000 mg/l

Denver, Colorado
  Annual evaporation rate = 29 in./yr
  Annual rainfall rate = 13 in./yr

Pueblo, Colorado
  Annual evaporation rate = 36 in./yr
  Annual rainfall rate = 12 in./yr

Lining cost:

30 mils (PVC) = $0.1322/ft^2, based on 60¢/lb

Fixed charge rate = .05 and 0.10

Land cost = $250/acre

Dike cost = $1.00/yd^3 (including equipment, labor and material)

Liner cover = $0.4/yd^3 (including equipment, labor and material)

Stripping land cost = $100/acre

Basis for installed pipe cost = $10.431/ft (for 12" diameter pipe)

Power cost = 12 mills/kwh

Pumps and accessories cost = $207.5/hp

Distance of conveyance = 100 miles

The solar evaporation costs $/Kgal. for various liner thicknesses are shown in Table 15.

TABLE 15
UNIT DISPOSAL COST BY SOLAR EVAPORATION

DENVER COLORADO

Brine Quantity mgd	Fixed Charge Rate,%	Unlined Ponds $/Kgal.	Lined Ponds (30 Mils) $/Kgal.	Lined Ponds (20 Mils) $/Kgal.	Lined Ponds (10 Mils) $/Kgal.
0.1	10	0.222	1.26	0.966	0.678
	5	0.111	0.63	0.48	0.34
1.0	10	0.108	1.125	0.83	0.54
	5	0.054	0.56	0.41	0.27
10.0	10	0.072	1.08	0.786	0.498
	5	0.036	0.54	0.39	0.25

*PUEBLO, COLORADO

Brine Quantity mgd	Fixed Charge Rate,%	Unlined Ponds $/Kgal.	Lined Ponds (30 Mils) $/Kgal.	Lined Ponds (20 Mils) $/Kgal.	Lined Ponds (10 Mils) $/Kgal.
0.1	10	0.194	1.035	0.796	0.564
	5	0.097	0.52	0.40	0.28
1.0	10	0.091	0.912	0.674	0.442
	5	0.045	0.46	0.337	0.22
10.0	10	0.059	0.874	0.635	0.403
	5	0.029	0.43	0.32	0.20

*Add 100 miles (12 Mills/Kwhr) pipeline cost from Denver, excluding right-of-way:

Brine Quantity mgd	Fixed Charge Rate, %	Pipeline Cost $/Kgal.
0.1	10	3.15
	5	1.81
1.0	10	1.25
	5	0.69
10.0	10	0.48
	5	0.26

Combined Process - Forced and Solar Evaporation

Creating a very large artificial salt lake bed near Denver can be objectionable from aesthetic and many other viewpoints. The best alternative for disposal of waste brine in this event is the combined process of forced and solar evaporation.

Preconcentration of waste brine to 10% solids by multistage flash evaporators will reduce the volume of waste brine by 93%, with the evaporator blowdown going to much smaller solar ponds east of the City of Denver which could conceivably be used as recreational salt lakes.

The cost presented for this process is based upon the assumption of the following:

Waste brine concentration = 7,000 mg/l to evaporators
Fixed charge rate = .05 and 0.10
Power cost = \$.025/Kgal. (based on L.P. steam turbine drives)
Steam cost = \$0.46/mbtu
Chemicals cost = \$0.016/Kgal. (cost of neutralization to pH 7.0)

(Note: Solar pond assumptions are based on Denver's annual evaporation rate.)

The forced and solar evaporation cost \$/Kgal. at 5% and 10% fixed charge rate and liner thickness are shown in Table 16.

TABLE 16

UNIT COST OF DISPOSAL FOR FORCED AND SOLAR EVAPORATION

DENVER COLORADO

Brine Quantity mgd	Fixed Charge Rate,%	Cost of Lined Ponds(30 Mils) \$/Kgal	Cost of Lined Ponds(20 Mils) \$/Kgal.	Cost of Lined Ponds(10 Mils) \$/Kgal.	Cost of Forced Evaporation \$/Kgal.
0.1	10	0.169	0.14	0.11	0.906
	5	0.085	0.07	0.055	0.661
1.0	10	0.126	0.096	0.068	0.715
	5	0.063	0.048	0.034	0.55
10.0	10	0.112	0.083	0.054	0.578
	5	0.056	0.041	0.027	0.454

CONCLUSIONS AND RECOMMENDATIONS

This study reveals that for Denver, Colorado, the lined solar pond is the best method for waste brine disposal at the present time. Ultimate disposal by deep well injection into the pre-cambrian breccia at 12,000 feet (which otherwise would be the cheapest method of disposal), is prohibited because of earthquake problems. Pipelining of brines to the ocean or the great salt lakes is, of course, prohibitively expensive.

Based upon ultimate disposal of 10 mgd waste brine at 7000 mg/l, using fixed charge rate of 7%, the following are recommended:

Solar evaporation east of Denver, if land can be found, would cost between 35¢ and 76¢/Kgal. with liner thickness of 10 and 30 mils.

In case sufficient land for solar ponds east of the proposed water renovation plant site at Denver cannot be procured, the following alternatives can be used:

> Running a pipeline to Pueblo, picking up brines from Colorado Springs on the way and putting solar evaporation ponds to the east of Pueblo in a fairly dry desert area. Solar evaporation east of Pueblo, would cost between 4.1¢ and 61¢/Kgal. evaporated, plus pipeline costs per 100 miles of 35¢/Kgal., a total range of 39¢ to 96¢/Kgal., depending whether or not liners are used.
>
> Forced evaporation of waste brines to 10% solids by multistage flash would cost 50¢/Kgal. Disposal of evaporator blowdown to solar ponds would cost between 4¢ and 8¢ with liner thickness of 10 to 30 mils. Thus the total for this combined process varies from 54¢ to 58¢/Kgal.

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NOMENCLATURE

A	Cost of 12" diameter pipe $ per linear foot (installed)
AE	Surface area; acres
ALNR	Area of liner; square feet
AWT	Advanced waste treatment
B	Cost of pump and drive; $/hp
B*	Cost of pump and drive, including 50% standby capacity plus 100% diesel standby generation; $/Hp
BTU;btu	British thermal unit
CE	Capital cost of solar evaporation pond
CE_u	Unit cost of brine disposal by solar evaporation; $/Kgal.
CFH	Cubic feet per hour
CLND	Cost of land; $/acre
CLNR	Cost of liner; $\$/ft^2$
CP_u	Unit cost of brine disposal by pipeline conveyance for ℓ miles; $/Kgal.
CT_u	Total unit cost of brine disposal by solar evaporation and pipeline conveyance for ℓ miles; $/Kgal.
D	Inside diameter; ft
ER	Net annual evaporation rate, in./yr
F	Evaporation pond fetch; miles
FCR	Fixed charge rate (annual capital recovery factor, fraction or % as needed)
FCR'	Modified fixed charge rate (including annual maintenance, fraction FCR + 0.0025)
f_d	Darcy's friction factor, dimensionless
gpd	Gallons per day
HW	Height of wave; ft
H	Effective height of formation face; ft
HP	Horsepower
h_o	Heat transfer coefficient, BTU/hr - sq ft °F
HD	Height of dike; ft
Kgal	Kilo gallons
Kwhr	Kilowatt hour
L	Length of dike; yards
ℓ	Pipeline distance; miles
mbtu	Million british thermal units

NOMENCLATURE (Continued)

MGD;mgd	Million gallons per day
mg/1	Milligrams per liter
MSFE	Multi-stage flash evaporator
N	Number of stages
OMR	Operation, maintenance, and repairs
P	Average formation porosity; volume of voids/total volume
P_d	Precipitate depth; ft
P_f	Pressure drop due to pipe friction; psig
psia	Pounds per square inch, absolute
psig	Pounds per square inch, gauge
P_t	Feet of precipitate per foot of water evaporated
PVC	Polyvinyl chloride
P_w^o	Operating well head pressure; psig
P_w	Well head pressure excluding friction; psig
Q	Brine disposal capacity; mgd
Q_e	Brine disposal capacity evaporated; mgd
Q_p	Brine disposal capacity for pipeline conveyance; mgd
R	Cost of right of way: $/mile
Rp	Optimum performance ratio
S_r	Brine disposal concentration ratio; S_p/S (dimensionless)
S	Brine disposal concentration from AWT treatment plants, 7000 mg/1
S_p	Brine disposal concentration in pipelining, mg/1
SCE	Submerged combustion evaporator
T_a	Total land area; acres
TDS;tds	Total dissolved solids
V_D	Volume of dike; cubic yards
V_f	Volume of fill to cover liner, cubic yard
Z	Power unit cost, mills/Kwhr

APPENDIX

It has not been practical to include in this report the details of all calculations carried out under the contract. Certain material has been collected together for possible review by those who have need of the greater detail. The material includes:

1. A more complete list of source documents

2. Mathematics and/or computer write-ups for:

    a. deep-well injection
    b. solar evaporation
    c. combined solar evaporation and pipeline conveyance
    d. pipeline conveyance
    e. optimization of multistage flash evaporators
    f. multistage flash evaporation with ion-exchange pretreatment
    g. vapor compression cost versus cost of power

This information, entitled "Supplementary Material under Contract 14-12-492 with Burns and Roe, Inc., Disposal of Brines, etc." is deposited in the Library of the Robert A. Taft Water Research Center, Cincinnati, Ohio, and may be examined there.

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